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PB-292 095

Proceedings: Symposium on New Concepts for Fine Particle Control

Notre Dame Univ, IN

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, NC

Aug 78

United States
Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

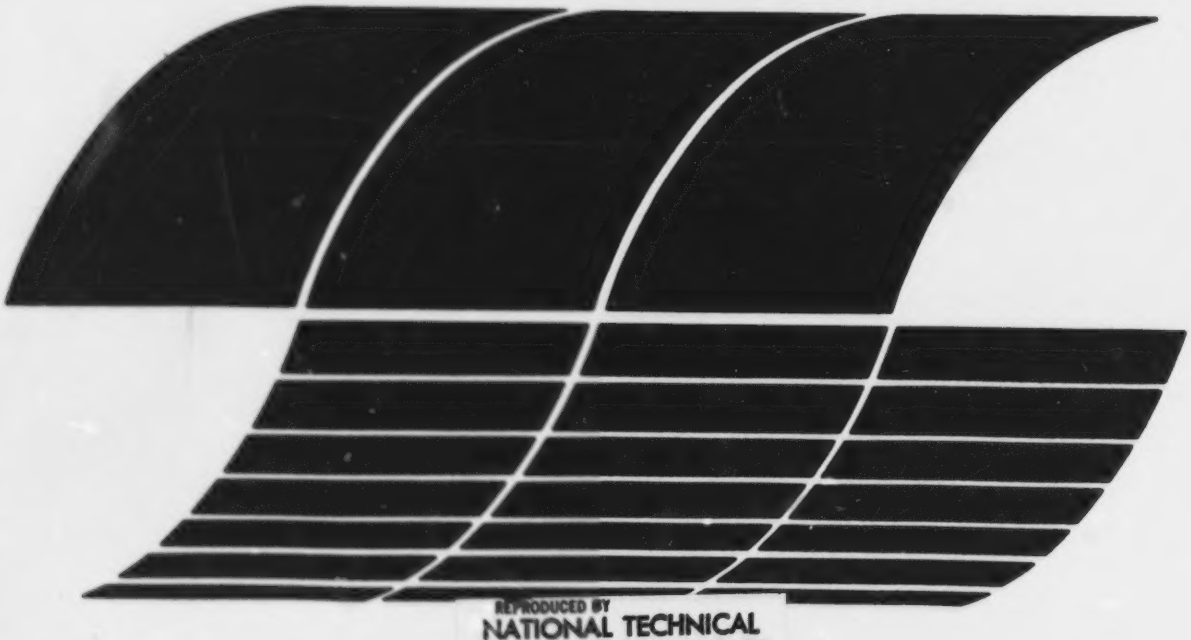
EPA-600/7-78-170
August 1978

PB 292095

EPA

Proceedings: Symposium on New Concepts for Fine Particle Control

Interagency Energy/Environment R&D Program Report



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TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/7-78-170
2.
3. RECIPIENT'S ACCESSION NO.

4. TITLE AND SUBTITLE Proceedings: Symposium on New Concepts for Fine Particle Control

5. REPORT DATE August 1978
6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S) Teoman Ariman, Compiler

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS
University of Notre Dame
Notre Dame, Indiana 46556

10. PROGRAM ELEMENT NO. EHE624

11. CONTRACT/GRANT NO. EPA Grant R805148; NSF Grant ENG. 77-02016

12. SPONSORING AGENCY NAME AND ADDRESS
EPA, Office of Research and Development(*)
Industrial Environmental Research Laboratory
Research Triangle Park, NC 27711

13. TYPE OF REPORT AND PERIOD COVERED Proceedings; 1/77-7/78

14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES Project officers are: Dennis C. Drehmel (EPA), Mail Drop 61, 919/541-2925, and Morris S. Ojalvo (NSF). (*) The National Science Foundation is co-sponsor.

16. ABSTRACT The report documents presentations made during a symposium on novel concepts, methods, and advanced technology in particulate/gas separation. The symposium, held at the University of Notre Dame and sponsored by the National Science Foundation and the Environmental Protection Agency, was held both to identify new research areas and to stimulate future research activities. The presentations included two general lectures by internationally known scientists (Dr. M. Benarie of France and Professor F. Loffler of West Germany) and 17 invited lectures by other prominent experts.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Pollution Control	13B
Dust	Stationary Sources	11G
Flue Gases	Particulate	21B
Measurement		14B
Separation		
Air Cleaners		13A

18. DISTRIBUTION STATEMENT Unlimited
19. SECURITY CLASS (This Report) Unclassified
20. SECURITY CLASS (This page) Unclassified
21. NO. OF PAGES 464
22. PRICE PC A20/MF A01

EPA Form 2220-1 (9-73)

EPA-600/7-78-170

August 1978

Proceedings:
Symposium on New Concepts for Fine Particle Control

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University of Notre Dame
Notre Dame, Indiana 46556

Grants: R805148 (EPA) and ENG. 77-02016 (NSF)
Program Element No. EHE624

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U.S. ENVIRONMENTAL PROTECTION AGENCY
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WORKSHOP PROGRAM

WEDNESDAY, APRIL 20, 1977

8:00 am Registration

8:30 am Welcome
JOSEPH C. HOGAN
Dean, College of Engineering
University of Notre Dame
Notre Dame, Indiana

8:40 am Opening Remarks
TEOMAN ARIMAN
University of Notre Dame
Notre Dame, Indiana

9:00 am Acoustics in Particulate-Gas Separation
Chairman:
K. T. YANG
University of Notre Dame
Notre Dame, Indiana

"Pulse-Jet Acoustic Dust Conditioning and Some New Potential Applications"

DAVID S. SCOTT
University of Toronto
Toronto, Canada

"Acoustic and Turbulent Agglomeration of Sodium Aerosols - Experimental Results"

WILLIAM C. HINDS and EUGENE F. MALLOVE
Harvard University
Cambridge, Massachusetts

10:30 am Coffee Break

10:45 am "New Applications of Acoustic Agglomeration in Particulate Emission Control"

DAVID T. SHAW and JAMES WEGRZYN
State University of New York at Buffalo
Buffalo, New York

"Effects of Electric and Acoustic Fields on Particle Collission Rates in Aerocollodial Suspensions"

PAUL D. SCHOLZ
The University of Iowa
Iowa City, Iowa

12:15 pm Lunch

1:30 pm High Gradient Magnetic Field in Particulate-Gas Separation

Chairman:
Benjamin Y. H. Liu
University of Minnesota
Minneapolis, Minnesota

"Engineering Aspects of Dry Magnetic Separation"

ROBIN R. ODER
Research and Engineering
Bechtel Corporation
San Francisco, California

"Magnetic Separation of Particulate Air Pollutants"

DENNIS C. DREHMEL
Environmental Protection Agency
and
C. H. GOODING
Research Triangle Institute
Research Triangle Park, North Carolina

3:00 pm Refreshments Break

3:15 pm "Research Needs and Opportunities in High Gradient Magnetic Separation of Particulate Gas Systems"

Y. A. LIU and C. G. LIN
Auburn University
Auburn, Alabama

"High Gradient Magnetic Separation: Review of Single Wire Models:

F. J. FRIEDLANDER
Purdue University
West Lafayette, Indiana

THURSDAY, APRIL 21, 1977

8:00 am General Lecture

Chairman:
MORRIS S. OJALVO
National Science Foundation
Washington, D. C.

"The Influence of Electrostatic Forces for Particle Collection in Fibrous Filters"

FRIEDRICH LÖFFLER
University of Karlsruhe
Karlsruhe, W. Germany

9:00 am Electrostatic Filtration

"Electrofluidized Beds for Industrial-Scale Air Pollution Control"

JAMES R. MELCHER
Massachusetts Institute of Technology
Cambridge, Massachusetts

"Approximate Equations for Predicting Electrostatic Particle Collection:

DOUGLAS W. COOPER
Harvard University
Cambridge, Massachusetts

10:30 am Coffee Break

10:45 am "Electrostatic Filtration and the Apitron-Design and Field Performance"

DENNIS J. HELFRITCH
Apitron
Buffalo, New York

and

TEOMAN ARIMAN
University of Notre Dame
Notre Dame, Indiana

"Electrostatic Capture of Fine Particles in Fiber Beds"

DONALD L. REID
Battelle Northwest
Richland, Washington

12:15 pm Lunch

1:30 pm All Other New Methods

Chairman
DENNIS C. DREHMEL
Environmental Protection Agency
Research Triangle Park, North Carolina

"Fine Particulate and Gaseous Emissions Control Experience with the TRW Charged Droplet Scrubber"

R. R. KOPPANG
TRW, Inc.
Redondo Beach, California

"Research on New Equipment for Cotton Dust Collection"

ALBERT BARIL, JR

DEVRON P. THIBODEAUX
Southern Regional Research Center
Department of Agriculture
New Orleans, Louisiana

and

ROBERT B. REIF
Battelle Columbus Laboratories
Columbus, Ohio

3:00 pm Refreshments Break

3:15 pm "Dielectrophoretic Air Filtration - Progress and Problems"

GEORGE H. FIELDING
Chemistry Division
Naval Research Laboratory
Washington, D. C.

"Fine Particulate Control Using Foam Scrubbing"

TOM E. CTVRTNICEK
Monsanto Research Corporation
Dayton, Ohio

"The Anomalous Behavior of Asbestos Fibers"

JAMES W. GENTRY
University of Maryland
College Park, Maryland

6:00 pm Social Hour

7:00 pm Banquet

Speaker:

TOM PAGNA
South Bend, Indiana

FRIDAY, APRIL 22, 1977

8:30 am General Lecture

Chairman:

ANDREW R. McFARLAND
Texas A&M University
College Station, Texas

"High Temperature Filtration - Technical Prospects and as Means for Check on Theories"

MICHEL M. BENARIE
National Institute for Applied Chemistry
Paris, France

9:30 am Panel Discussion

"New Directions in Particulate-Gas Separation Research"

Moderator:

BENJAMIN Y. H. LIU
University of Minnesota
Minneapolis, Minnesota

Panel Members:

TEOMAN ARIMAN
MICHAEL M. BENARIE
DENNIS C. DREHMEL
ROBIN R. ODER
DAVID S. SCOTT

10:30 am Coffee Break

10:45 am Continuation of the Panel Discussion

12:15 pm Closing Remarks

TEOMAN ARIMAN

12:30 pm Lunch

1:45-2:30 pm Guided Tour of the Campus

ACKNOWLEDGEMENT

The editor of this proceedings gratefully acknowledges at every phase of the Workshop, the interest and continuous support and advise of Dr. Morris S. Ojalvo, Director of the Particulate and Multiphase Processes Program of the National Science Foundation and Dr. Dennis C. Drehmel of the Industrial Environmental Research Laboratory of the Environmental Protection Agency. Special thanks are due to Mrs. Shirley Wills for her enormous efforts in the transcription of the taped sessions of the workshop and her expert typing of most of the proceedings. I would also like to acknowledge the diligent work by two of our students at Notre Dame, Mr. Greg Muleski and Mr. Carl Aumen in the preparation of the open discussions and panel session and in editing of the proceedings.

WELCOME ADDRESS

Joseph C. Hogan
Dean, College of Engineering
University of Notre Dame
Notre Dame, Indiana 46556

Let me welcome you to the University of Notre Dame. I'll not go through the Chamber of Commerce sales pitch because engineering schools are known by the faculty they have and I hope you are familiar with members of our faculty. You are more than welcome to go over to the Engineering Building, but you will have to tread through a good deal of mud to get there. This is due to the eight million dollar addition which is just being started and should be completed by December of next year. If any of you would like to see that particular construction, I'm sure K. T. or Teo will be glad to make arrangements for this or anything else you'd care to see while you're here.

We are honored to have such a distinguished group here. It is farsighted for NSF and EPA to act as co-sponsors. Often a field moves forward in tiny increments because the papers and work are presented at conferences at random times. I think that getting the top people in the field together at one time is without question the best way to move the field forward. I want to congratulate NSF and EPA again for their farsightedness and for sponsoring this program. I hope it will prove to be an interesting and productive two and a half days. Thank you.

OPENING ADDRESS

Teoman Ariman
Workshop Director
Department of Aerospace and Mechanical Engineering
University of Notre Dame
Notre Dame, Indiana 46556

Welcome to the Workshop entitled "Novel Concepts, Methods and Advanced Technology in Particulate-Gas Separation." As the Dean said, the workshop is sponsored by NSF and EPA and we would like to express our appreciation, especially to Dr. Morris Ojalvo of NSF and Dr. Dennis Drehmel of EPA, for help with the workshop and continuous support. We would also like to thank the members of the planning committee, Professor Benjamin Liu of the University of Minnesota and Professor K. T. Yang, Chairman of the Aerospace and Mechanical Engineering Department here at Notre Dame. We sincerely hope that everything went well last night and that the transportation and hotel accommodations are satisfactory. If there are any complaints, please see Dr. Yang or Dean Hogan. If there are any compliments, please see me.

I would like to briefly introduce the workshop. As you know, a number of emission control systems have been used for years--fabric filters, electrostatic scrubbers, electrostatic precipitators, etc. In years past, energy has been quite cheap. But times have changed and we face an energy crisis unless we do something. Furthermore, we face the challenge of controlling fine particulate emissions. Taken together, these two demand that we search out new technologies, and in the past few years, there has been substantial growth in this area. The thrust of this workshop is the bringing together of people and their technologies. We wish to discuss not only collection efficiency but also energy efficiency. With the close work of the planning committee and the NSF and EPA, it was decided to invite three or four people in each area - a theoretical analyst, an experimentalist and a design/applications specialist. By mixing people from universities, government laboratories, private laboratories, and industry, we hope to have a healthy exchange of ideas. I sincerely hope that the workshop reaches that goal.

STATE OF THE ART AND OBJECTIVES OF THE WORKSHOP

Teoman Ariman
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Department of Aerospace and Mechanical Engineering
University of Notre Dame
Notre Dame, Indiana 46556

Traditionally, developments in particulate control equipment have been slow in starting, as evidenced by the fact that today's basic equipment differs little from that in use 100 years ago. One major reason for the lack of progress is that successful innovation requires field trials, where the cost of failure to the trial plant is large. Therefore, the risk must be small, and innovative can be only a small departure from the traditional design. Moreover, most industrial applications of air pollution control equipment are considered as non-productive economic burdens on the processes to which they are applied. As a result, control devices so far has received minimum attention, and incentive for research and development to improve their performance is lacking.

Fabric filtration systems have been employed in industry for over a century with relatively few technological modifications [1,2]. Indeed, there have been in recent years only a few advances of note: introduction of fabrics capable of withstanding higher temperatures and development of pneumatic cleaning techniques. For the most part, there was in the past little incentive for improvement. The collection efficiency of the systems was mostly satisfactory and the units were inexpensive to purchase. However, with the recent substantial increase in energy costs attention should now be given to operating expenses (e.g., energy consumption). As a result, the filtration systems of yesteryear may not be the best approach for future applications.

Recently a new method has been developed for the particulate-gas separation through the fabric filtration. Some initial analytical and/or experimental investigations have been performed (e.g., Zebel [3,4], Butterworth [5], Ariman [6-9] and Penney [10]) for the particulate-gas separation by fiber and fabric filters in an electrostatic field. It has been observed that the application of an external electrostatic field improves the efficiency of fine particle collection considerably. More recently Ariman has begun to investigate the pressure drop across a flat fabric filter due to the collection of electrically charged dust particles. There was no electrostatic field placed across the filter. Two industrial dust samples, fly ash and limestone, and Nomex 14 oz.fabric filter samples were utilized in the bench scale tests. It was found that under the same operating conditions the pressure drop across the dust layer and filter was reduced by 1/2 to 2/3 as compared to the conventional noncharged dust particles case. An electron-microscopic study of the dust cakes for charged and noncharged cases indicate that charged dust particles deposit in a relatively porous layer resulting in a low pressure drop across the filter. The resulting more porous dust layer also facilitates cleaning of the filter, since it tends to fall off readily, leaving the cloth relatively clean. It was also found that there was considerable improvement in the collection efficiency.

It appears that this new method has a great potential in particulate-gas separation by fabric filters and may bring enormous changes in the conventional filtration area. However, the research performed to date in this field has only just scratched the surface and a number of unknown aspects of this highly promising area need immediate attention of researchers [12-34]. For example the electrical behavior of aerosols, electrostatic charge on fiber and fabric filters, particulate cake structure, collection efficiency and pressure drop and a number of other unknown aspects of the field need to be investigated.

Another new method in the particulate-gas separation has recently been initiated by Drehmel and Gooding [35]. The method is directed toward a combination of magnetic and filtration collection mechanisms. In its simplest practical form, the high gradient magnetic separator (HGMS) consists of a canister packed with the fibers of a ferromagnetic material (such as stainless steel wool) and magnetized by a strong external magnetic field [35]. The strong magnetic forces produced near the edges of the fibers are cabable of trapping fine, weakly magnetic particles with high efficiency.

The packing density of the separation zone can be adjusted to obtain a proper balance between conventional filtration phenomena and the HGMS effect. Reducing the packing density lowers the pressure drop through the separator but requires a higher magnetic field for efficient particle collection. The overall particle collection efficiency is theoretically a function of the applied magnetic field, or filter mesh parameters (fiber diameter and magnetization, packing density, and length of mesh in direction of flow), of particle parameters (magnetic susceptibility and diameter) and of fluid parameters (superficial velocity and viscosity).

To date the high gradient magnetic separation has been applied extensively to particle liquid separations [36-39]. The bench-scale experimental work performed by Drehmel and Gooding [35] appears to be the first known attempt to apply HGMS to particle-gas separations. It is known that conventional magnetic separation is generally useful for the separation of relatively large particles of strongly magnetic materials such as iron and magnetite. HGMS, by contrast, makes possible the efficient separation of micron-sized particles of weakly paramagnetic materials from liquids at high process rates. It can be shown through generalized theory that application can also be made to particle removal from gas streams. Magnetic forces can be cost effective if the control device is properly designed. One of the most important feature of the recent work by Drehmel and Gooding [35] is the dramatic improvement of collection efficiency when a low-strength magnetic field was applied. However, at present substantial analytical and experimental research efforts are needed to identify other particle-gas separation problems to which high gradient magnetic separation might be applicable, to develop analytical models based on the experimental investigations and to provide economic estimates for specific promising applications such as particulate emission control from basic oxygen surfaces.

Another recently developed new method in particulate-gas separation is based on acoustic coagulation. Recently there has been a substantial growth in research dealing with various aspects of acoustic-aerosol interactions. In the last few years considerable attention has been devoted to the area of acoustic propagation and attenuation in aerosols [45]. Theories have been developed for both small amplitude [41-42] and large amplitude [43-44] sound waves, including the processes of mass, momentum and energy transfer between droplets and the surrounding gas-vapor mixture. It is well known that acoustic streaming can substantially increase the rates of convective transport of mass and energy and chemical reactions at solid-fluid and fluid-fluid interfaces, provided the sound pressure level is sufficiently high [45]. The aforementioned investigations have indicated [46] that the acoustic streaming mechanism may be of importance in droplet evaporation, particle drying, acoustic coagulation, and droplet or particle combustion, especially in those cases where the droplets or particles are large enough so that the acoustic boundary layer thickness is much smaller than the droplet radius. In that case very significant increases in heat and mass transfer are possible [47].

Initial experimental investigations were directed to the use of shock waves to coagulate fine particles [48-50]. They also included the use of pulsating combustion burners for providing the shock or more correctly saw-tooth waves, i.e., sound waves with periodic discontinuities for agglomeration [51]. The pulse output approximating a saw-tooth wave with a step wave front has successfully agglomerated ZnO_2 aerosols to bring about a five-fold increase in mean particle size [51]. Since no special agglomeration chamber was used, the experiment was equivalent to placing the pulse-jet in a suitable elbow of the duct transporting the dusty gas in an industrial plant. Recently Scott and Rennick have experimentally investigated several aspects of acoustic-aerosol interactions in finite-amplitude, mutually perpendicular, standing and progressive acoustic fields [52-55]. Acoustically generated turbulence and the role it may play in aerosol agglomeration has been more recently investigated [55-56].

Another recent experimental program has been initiated by Shaw [57] and aims to provide information on the experimental investigation of acoustic coagulation in the flowing coagulation chamber, the evaluation of sinusoidal standing wave acoustic system versus saw-tooth traveling wave acoustic system and on the search and testing of large acoustic generators.

As indicated here although in recent years there has been a substantial growth in research dealing with various aspects of acoustic-aerosol interaction, a number of vital investigation areas need to be identified in this promising field.

New concepts and methods in the particulate-gas separation are not restricted to the stated three major areas. For example in the recently developed Charged-Droplet Scrubber (CDS), the scrubbing droplets, usually water, carry a high electrical charge which is induced during or after formation [58-59]. The droplets may move under the influence of electric fields, and the particulate may also be electrically charged. More investigation may be needed for the various charged droplet scrubbing mechanisms toward the control of fine particulate (0.1 - 1.0 μm in diameter).

In another research effort, an experimental investigation has been conducted by Scholz [60] on the effects of sonic and electric fields on the collision rate between suspended submicron sized particles. Recently a number of new concepts have been considered for the collection of submicron particles, specifically cotton dust in the Southern Region Research Center of the Department of Agriculture [61]. However, these developments are at the initial stage and a number of unknown aspects of these efforts may need immediate attention of investigators.

As indicated above, all three new methods and the other new techniques appear in the urgent need of a number of investigations. In order to identify new research areas associated with these methods as well as to stimulate future research activities in these fields, the research workshop on the novel concepts, methods and advanced technology in particulate-gas separation was held with the sponsorship of the National Science Foundation and Environmental Protection Agency at the University of Notre Dame.

A technical committee consisting of Drs. T. Ariman, B. Y. H. Liu and K. T. Yang, was in charge of the technical content of the research workshop. Drs. Ariman and Yang have been associate professor and professor and chairman respectively at the Department of Aerospace and Mechanical Engineering of the University of Notre Dame, while Dr. Liu has been a professor of the Department of Mechanical Engineering and the director of the Particle Technology Laboratory at the University of Minnesota.

Three groups of researchers were invited by the technical committee; 1. theoretical analysts; 2. experimentalists; 3. design and application specialists. The interaction between the three groups through presentations and intensive discussions is emphasized particularly, because very often a researcher in one group can benefit from the other groups. It was quite gratifying that practically all participants attended the workshop in its entirety and took an active role in every aspect of it.

The workshop took place on April 20, 21 and 22, 1977 at the University of Notre Dame Campus. The number of participants was limited to 56 scientists and engineers from universities, government and other research organizations, and industry. Two general lectures by internationally known scientists, one by Dr. M. Benarie of France and another by Professor F. Löffler of W. Germany, and 17 invited lectures by the prominent experts were presented. The time allowed for general and invited lectures were 45 and 25 minutes respectively. Each presentation was immediately followed by an in-depth discussion of 15 minutes for general and 20 minutes for invited lectures. The latter part generally began with a designated discusser for each presentation and then a general discussion open to every participant was conducted. Each designated discusser made a considerable preparation for a constructive in-depth discussion of each presentation and submitted his written discussion well before the workshop date. In order to stimulate the discussions, the copies of the manuscripts were distributed to the relevant discussers prior to the workshop. On the morning of Friday, April 20, a panel discussion on "New Directions in Particulate-Gas Separation Research" was conducted. The formal presentations by the members of the panel and concise discussions and comments by participants were quite productive in identifying new research topics in the particulate-gas separation area. The entire workshop sessions were tape-recorded and the tapes were extensively utilized in the preparation of these proceedings.

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PULSE-JET ACOUSTIC DUST CONDITIONING AND SOME NEW POTENTIAL APPLICATIONS

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I INTRODUCTION

Acoustic conditioning of fine-particle aerosols is a process by which the mean size of the particles is increased and their number density decreased through exposure to finite-amplitude acoustic fields. This change in the particulate size distribution is important – for it can allow an increase in the collection efficiency of a downstream dust collector and/or a reduction in the overall particulate separation cost because, with few exceptions, collecting devices retain higher mass fractions as the particulate size is increased. The increased coagulation rate which achieves this rapid change in the size distribution is the result of manyfold increase in the particle-particle collision frequency – which in turn results from dynamic acousto-aerosol interactions. Experimental evidence, and straightforward physical arguments, indicate that the process can be particularly effective on high dust load fine particle aerosols.

Although the principle of acoustic dust conditioning (ADC) has been known for sometime and has been shown to be technically effective, the process has enjoyed little industrial application. In my view, this lack of industrial application has been the consequence – primarily – of economic limitations. The first of these is high operating costs due to high specific power requirements, and the second is high capital costs.

Studies performed at the Ontario Research Foundation and the University of Toronto have examined a new approach to ADC – in which the following two complimentary principles were explored.

(i) Firstly, the conventional standing sinusoidal acoustic field, used in essentially all earlier acoustic dust conditioning work, was dropped in favour of progressive saw-tooth fields. This change in acoustic configuration allowed major simplifications in the agglomeration chamber yielding both capital and operating cost savings.

(ii) Secondly, the normal techniques of sound generation – such as sirens and mechanically activated pistons – were discarded in favour of a resonant pulse-jet acoustic source. This not only allowed a direct conversion of fuel to sound and the use of a simple no-moving-part device, but also readily produced high intensity progressive saw-tooth acoustic fields.

The results of this research/development program are thoroughly reviewed in Reference 1. In summary, the practical results of this program showed that significant – 0(10) – advances can likely be achieved by this new process in all three areas of (a) capital, (b) maintenance, and (c) operating costs. At the same time, the mean particle size was increased 5-7 fold. But as noted, the details of this research are well covered in Reference 1 and it would be inappropriate to review these here.

II THE NEXT STAGE OF ADC DEVELOPMENT

Since the work described in Reference 1 was performed, the acoustic-aerosol research with which I have been involved has primarily concentrated on more fundamental acousto-aerosol phenomena. Insofar as the applied aspects of pulse-jet acoustic dust conditioning we have, as they say in the performing arts, been between shows. And this gives us time to pause and take note of where we are. To date, it seems to me that research on pulse-jet dust conditioning has indicated economic and technical feasibility *in principle*. The next stage in the systematic development of these concepts must examine promising conditioner/collector *systems* in the context of likely applications. It is this next stage of the development, to which I address the remainder of this paper.

But by way of philosophical introduction, I preface my remarks by saying that it is this stage which is, perhaps, the most critical. The history of acoustic agglomeration has been marred by overzealous researchers who have felt the process to be a likely panacea for all fine particle collection requirements, as well as by ambitious entrepreneurs anxious for solution to specific problems on short-time-scale development programs. My own view is that ADC will never be a panacea-type solution to fine particle removal from industrial sources. But if well conceived development programs are carried out with selected conditioner/collector systems, and with specific applications in mind (and over development time scales that can allow careful and systematic evaluation of the relevant parameters) then it seems reasonable that in specific technical scenerios, ADC in combination with an appropriate collector, could be more effective than any alternate approach. Moreover, once we have identified *single* application in which this situation exists, the value of careful development will have been illustrated. And by that impetus the process development can be expanded to further applications.

But care – and perhaps some luck – is required in the choice of both the conditioner/collector configurations as well as the problem to which they will be applied. And those funding this development must not be mislead into thinking that conclusive results can be obtained in a matter of months.

With this background, I will briefly outline two development projects with which I am associated, and which have been initiated within the past few weeks. Clearly, I can only describe the work we hope to do rather than present any results.

2.1 ADC/Wet Scrubbers in the Pulp and Paper Industry

In Reference 1, wet scrubbers are identified as perhaps the most promising of collector systems for use in combination with ADC. To illustrate by example; in the case of a particular venturi scrubber, 3,150 bhp was required to collect 90% of the 0.2 μm particles while only 60 bhp was required to collect 90% of the 0.5 μm particles. The implication for ADC is that by affecting a 2.5 fold increase in mean particle size (we have previously noted 5~7 fold increases) a 50 fold *reduction* in scrubber power requirements would be realized. While this illustration refers to a particular venturi scrubber, all scrubbers exhibit sharp transitions in collection efficiency as a function of particulate size within the size range 0.1~1.0 μm.

Of course wet scrubbing can also be used to simultaneously remove noisome (noxious/odorous) gases. And there is some evidence that the presence of a finite-amplitude acoustic field can also improve noisome gas scrubbing effectiveness. In this circumstance, the acoustic field might be considered to act both as a *conditioner* for particulate collection and as a *catalyst* for noisome gas removal. Although the latter potential benefit is more speculative, it could be important in applications within the Pulp and Paper Industry and so has been included in our program.

A possible deleterious effect on scrubber performance would occur if the sound were to rapidly agglomerate the liquid droplets in the scrubber itself, thereby markedly reducing the surface volume ratio of the scavenging liquids. Whereas it is possible to visualize a situation in which this occurs – it is also apparent that configurations can be designed to minimize this effect if it should prove important, as proper design undoubtably involves some balance between maximizing the droplet "swept" volume and minimizing droplet agglomeration.

The research program will proceed as follows.

Phase I

(1) Visits will be made to various pulp and paper mills for the purpose of familiarization with the practice and potential of wet scrubber gas-cleaning. This study will include an appraisal of (a) existing scrubber installations, (b) potential new installations both in new mills and as replacements for existing non-scrubber equipment, (c) potential for pulse-jet conditioning of existing scrubbers for the purpose of upgraded performance and/or reduced annual costs, (d) identification of the mill environment in gas-cleaning locations for the purpose of evaluating noise isolation requirements.

(2) With the benefit of (1), the experimental program will be finalized. In particular, key parameters requiring examination will be identified and a desired data point grid constructed.

Phase II

(3) A parametric evaluation of the effectiveness of pulse-jet acoustic conditioning on particulate collection by a medium energy venturi scrubber will be performed. Flow rates between 1000-3000 cfm will be studied. A ZnO fume will be generated on-line and used as the test aerosol. By this means, particulate size distributions and densities can be readily achieved which properly model typical pulp and paper aerosol effluents.

(4) The influence of intense acoustic fields on gas scrubbing performance in the above facility and at flow rates of 1000-3000 cfm will be studied. The test noisome gas will be SO_2 and the test scrubbing liquid an Na_2CO_3 solution.

(5) If practical from an experimental point of view, the pulse-jet acoustic field will be subdivided into a dirty gas stream and a clean gas stream in order to obtain data for

extrapolating to the ultimate flow rates which may be treated by the pulse-jet. (It is expected that the pulse-jet will be able to condition a much greater flow volume than the 3000 cfm generated in this program).

Phase III

(6) Based upon the results of Phases I and II, an economic evaluation of the capital, maintenance and operating costs of pulse-jet conditioned wet scrubber systems will be performed. This will be structured in the form of annual gas-cleaning costs and compared with pure scrubber systems for specified collection efficiencies and aerosol characteristics. The economic evaluation will include an estimate of acoustic isolation costs for typical pulp and paper gas-cleaning environments. Annual gas-cleaning cost comparisons will be made with other collectors.

2.2 ADC/Mechanical Separators in High Pressure-Temperature FBC Applications

Development of fluidized-bed coal combustors for use in thermal power generating stations is being carried out at several laboratories. The process involves combusting pulverized coal in a fluidized-bed of either pulverized limestone ($CaCO_3$) or a synthetically prepared calcium-bearing particulate. The lime solids react with the combustion released sulphur oxides to form $CaSO_4$ which remains in the bed as particulate matter. Thus, the aerosol effluent contains a much lower level of SO_2 in the carrier gas – but the particulate mass loading can be much greater (due to entrainment of calcium particulate along with the ash particulate) – than is the case in conventional combustion processes. In a sense, the gaseous pollution problem is reduced at the expense of increased particulate loading.

In a fluidized-bed combustion (FBC) configuration being examined by Argonne National Laboratories, the combustion process is pressurized to approximately 10 atmospheres with the intention that the off-gases be expanded through gas turbines. But hot, particulate laden, gases cannot pass through gas turbines without massive destruction to the turbine blades resulting in unacceptably short lifetimes. And so we are faced with a particulate-gas separation problem at high temperatures and pressures that is critical not only for environmental considerations but also for the success of the process itself.

At the temperatures O(400°K) and pressures O(10 atm.) which are anticipated, most conventional particulate removal techniques are unsuitable. Mechanical, or inertial, separation by cyclones seems to be one of the few systems that can, in principle, operate at these temperatures and pressures. Unfortunately, the performance characteristics of cyclones are not sufficient to meet the gas cleanliness standards set by turbine manufacturers – although there remains some uncertainty regarding what these standards should be. It now appears that the ultimate criteria may require that there be no more than 0.5 mg/m^3 of particulate matter above 2 μm diameter.

Results of the FBC program at ANL have shown that the mass loading and size distribution of the particulate matter exiting from the second of two inertial separators is not far from that which would be acceptable as turbine inlet conditions. Moreover, a fairly straightforward evaluation indicates that if the size distribution of particulate matter to the entering the cyclone was increased by perhaps less than a factor of 5, the performance of the cylcone would be improved such that the effluent *would* meet turbine inlet criteria. For these reasons the group at ANL have decided to initiate a research program to investigate the effects of using ADC in this application. And it is in this regard that I have become associated with these activities.

The program at ANL will evaluate effectiveness of ADC in a bench scale fluidized-bed combustor. It is proposed that the sound source be a resonant pulse-jet and that progressive sawtooth waves constitute the acoustic field configuration.

There are several promising aspects of this approach.

For instance the heat of combustion from the pulse-jet simply adds to the overall process heat. Thus the effective power requirements will, in principle, be zero – although the cost per heating unit for pulse-jet fuel may be expected to be somewhat higher than that of coal. Furthermore, the pulse-jet will run at higher power levels due to the elevated pressures and should, in principle, be a more effective sound generator than if it were running at ambient conditions.

Conceivably, there might also be deleterious ADC effects. For instance, very fine submicron particles could be agglomerated into a size range below that which would be effectively removed by the cyclone, but within a range which could do damage to turbines. Evaluations of the mass loading ratio in the very fine particle spectrum indicate that this should not be a problem – but it will clearly be an aspect which shall be monitored during the experimental program.

Turning to the experimental program itself, one of the main difficulties in this initial stage of process evaluation shall be "scaling". Put simply, a pulse-jet is a device which cannot be scaled up or down without changing the acoustic characteristic – for a pulse-jet is a quarter wavelength device. Ultimately, it is expected that single pulse-jets will generate sufficient sound to treat between 5,000 to 10,000 cfm. But the current test plant at ANL yields approximately 81 scfm, or on the order of 20 acfm.

In response to these scaling demands, the current intention is to design a pulse-jet that will exhaust into a resonant-manifold. The manifold shall be capable of splitting the off-gases into a "waste" stream and two "process streams". Two process streams are required in order to examine the affects of ADC on both the primary and secondary cyclone performance. By means of the resonant-manifold a large total acoustic power output (Watts) can be reduced in total magnitidue to match the amount of flue gas being treated. At the same time, this approach will allow the acoustic field to be varied in intensity from approximately 165 db downwards, in order to parametrically evaluate sound intensity requirements. It is my feeling that by exhausting the pulse-jet into a resonant-manifold, and then drawing off the acoustic power as required for the much smaller scale dust treatment streams, we will both eliminate the scaling difficulties as well as allow a much greater degree of parametric flexibility, than would be the case if we exhaust the pulse-jet directly into the particulate laden gas. Of course the negative of this approach is that it requires the design, construction and debugging of a portable pulse-jet/resonant-manifold system.

Although the experimental program I have just described is still very much in the planning stage, and alternative programs may, in fact, be followed — the entire concept of applying ADC to high pressure, high temperature, particulate removal sytems is, I think, very exciting.

III CONCLUDING COMMENTS

It is my view that ADC has reached the stage where economic and technical feasibility have been shown to be feasible in principle. We must now carefully select certain applications where ADC may have clear advantages over alternate systems, and then perform careful unhurried evaluations of its capabilities in these technical scenerios. Perhaps one of the two applications I have reviewed in this paper — and which will be evaluated over the next 1~2 years — will be the application that gives us the first real illustration of an ADC/collector system meeting the required criteria better than any alternative process. I hope other investigators, and laboratories, seek other unique applications where their special requirements might yield an "application breakthrough".

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OPEN DISCUSSION

Billings: I think there is a misconception in accordance with what we might expect. We would all agree that agglomeration of some sort, whether mechanical, thermal, or turbulent, is desirable to improve the mechanical collector. I think zinc oxide, as an aerosol, is a questionable aerosol to use.

What you said was that you went down in concentration and up in apparent particle size. Correspondingly, the aerodynamic diameter parameter would go down. Although this may be 2.0 to 7.0 microns, the fact of the matter is that the momentum relaxation time is about 0.1 of the original value. This is what you should be measuring.

Scott: I agree with what you are saying. One thing that was not brought out, and for obvious reasons, is the kinematic viscosity of the gas. Aerodynamic diameter is a characteristic parameter of the aerosol and the momentum relaxation time involves both the characteristics of the gas and the particles.

I would like to clarify that all the sizes talked about are aerodynamic sizes and are a measure of the momentum relaxation time or of aerodynamic diameter. Results of size distribution were obtained from a cascade impactor. The means of measuring size is a means of predicting improved performance since collectors, particularly cyclone collectors, are inertia collectors.

Melcher: What are the resonance times requires for gas treatment to go from 0.5 to 2.0 microns?

Scott: Approximately between 1.5 and 2.5 seconds. Resonance times must be a factor of concentration. Acoustic vibration constants are inversely proportional to the concentration. So you should put the stuff in front of a conventional device where the concentration of particles is high. This is a necessary requirement for acoustic coagulation.

Scott: Is it possible to amend the resonance times by using an adverse density gradient?

Shaw: If you have an amount of the fine particles, your acoustic agglomeration is small so the count is large. With a large standard deviation you need large particles to act as collectors for the small particles. It is a complicated function of standard deviation, mass concentration and frequency.

That is why I am skeptical about the recent report by the EPA. It is based on the wrong equation, wrong data and, of course, has the wrong conclusion.

Another point I'd like to bring up is that because of economic comparison on acoustic agglomerators with conventional devices, comparisons can only be made for specific applications. A general statement is too dangerous. It can be too optimistic or too pessimistic.

Something else worthy of note is on the acoustic device, for instance, sirens; the last reference I could find was published in 1949. Bell Laboratories were to build a siren to put on the Empire State Building for air raid warnings. Acoustic devices must be studied using up-to-date information on acoustic agglomoration.

Finally, I have a question. By using a pulse generator, isn't a combustion product used which may get into the system?

Scott: To further elaborate on resonance times, no studies were done on parametric times for the resonance frequency. We are now studying an oil mist aerosol at the University of Toronto (UT) and have found that the change from uncoagulated to fully coagulated particles with no further change is very rapid. Well, under 1 second. The density was above 1 grain/ft^3. So we can assume that resonance times are not a problem.

From my work at Argonne National Laboratories, we found that by using a spiral configuration it would be easier to modify resonance times simply by changing the length of the spiral. Parametric evaluations would be easier to conduct.

From my work through the Ontario Research Foundation, we found that economics depend mainly on the costs attendant to acoustic isolation. Total capital, operating and maintenance costs ran $.50 to $1.50 per cubic foot. So it is economically feasible as compared to current devices, but it is a cost which must be added on since it is a conditioning. It is within or under costs of current devices.

Concerning the pulse jet, the products of combustion did go into the gas stream. This represented less than 1% of the total. In doing preliminary design work at Argonne on resonant manifolds, I think it will be possible to drop the products of combustion which get into the gas stream down to zero. A resonant manifold system can be used in almost any application.

Koppang: What is the actual energy input in watts/ft^3?

Scott: I don't recall right off hand, but it is in the paper.

Baril: Because there are maximum noise levels in the U.S., we must be concerned that the noise from the system isn't too loud. Is this a problem?

Scott: Not really. The pressure levels achieved are not larger than those from an internal combustion engine. Since mufflers can be built for cars, it should be no problem to build one for the system. This will be even easier since an internal combustion engine operates at a range of pressure levels, whereas the acoustic device has a resonant frequency and the muffler can be tuned to that frequency.

B. Liu: The agglomeration that you observed at a bend was irrespective of the pipe. That would mean that the ortho-kinetic agglomeration is really not very effective. Rather it is based on something that is acted upon through turbulence and the acoustics.

Scott: Your explanation is very good. Ortho-kinetic interaction is importand because you can describe it physically and everyone can believe it. I don't think it is the important mechanism.

One of the things that has come out of our research at UT is that an acoustic field itself can generate a turbulent field, even if no turbulence existed (we think). If you have a turbulent field with a flow process, then it is clear that an acoustic field adds to that turbulence. I think there is a whole set of messy processes occurring here that increases particle-to-particle collisions. And that's where I'd like to stop until we get some more fundamental work in.

It is interesting, diagnostically, to look at this question though. Because we are now using microfilms, hot wires and LD-8's (which go very well with the fact that we are using particles) the diagnostic procedures should improve the information gathered.

ACOUSTIC AND TURBULENT AGGLOMERATION OF SODIUM AEROSOLS

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ABSTRACT

Tests of turbulent and acoustic agglomeration of captive sodium fire aerosols at concentrations of 0.1 to 20 gm/m^3 were conducted in 90 m^3 and 0.65 m^3 vessels to evaluate these mechanisms for direct application air cleaning systems. Aerosol mass concentration decay with time was monitored by sequential filter samples. Turbulence was generated mechanically with a 51 cm diameter centrifugal fan impeller and a reverberant acoustic field was created with an electronic siren. The effectiveness of each method over a range of particle concentrations and power densities was evaluated by an agglomeration index, a measurement of particle growth based on sedimentation characteristics. Both turbulent and acoustic treatment markedly enhanced sedimentation rate compared to undisturbed settling. The effectiveness of both methods increased with increasing aersol mass concentration and increasing power input per unit volume of aerosol. The agglomeration index reached 20 for turbulent agglomeration at an aerosol mass concentration of 3 gm/m^3 and 7 for acoustic agglomeration at 14 gm/m^3 when using an acoustic intensity of 145 dB. Turbulent agglomeration was more effective than acoustic agglomeration for the same mass concentration and power density conditions.

INTRODUCTION

A study is being conducted in the Harvard Air Cleaning Laboratory to examine the feasibility of using direct application air cleaning as a treatment method for high concentrations of sodium fire aerosol that would likely fill a reactor containment vessel following a release inside a liquid metal fast breeder reactor. Most particulate air cleaning systems are dynamic, flow-through devices where particulate matter is continuously removed from a stream of dust laden gas as it passes through the device. A less common, approach can be applied when the dust laden gas is confined within a vessel. This approach treats the entire gas volume simultaneously, as a batch process, to reduce airborne particulate concentrations as rapidly as possible.

Leakage from the containment vessel is expected to be small, (no more than 0.1% of the containment volume per day). Because it is continuous, significant public health benefits to the surrounding community would accrue from a rapid reduction in airborne particulate concentration within the containment vessel. To reach the design goal, reducing by a factor of 10 the two-hour integrated concentration, it is necessary to reduce the airborne concentration within a matter of minutes by 90% or more. Although rapid clean up can be accomplished by a conventional air cleaning loop, such a system would have to be very large and expensive to treat a containment volume of $7x10^4$ m^3 in a few minutes. A direct application air cleaning approach, on the other hand, has considerable potential for this situation because it treats the entire containment volume simultaneously without the necessity of having to pass large volumes of gas through air cleaning devices.

Three approaches have been evaluated experimentally: scavenging by inert powder injection, enhancement of coagulation by mechanically generated turbulence, and enhancement of coagulation by high intensity sound. The latter two methods are similar and are the subject of this report. Inert powder injections have been described in Reference 3. Both turbulence and high intensity sound enhanced coagulation by applying energy to the gas, throughout its volume, to create relative motion between particles.

Because of the complex geometry of a containment vessel, including its machinery, the use of a tuned sonic coagulation system was not considered feasible and consequently no attempt was made to evaluate such a system. Instead, random traveling waves in a reverberant enclosure were used. Mechanically generated turbulence was produced with a centrifugal blower wheel. The experiments provided some insight into coagulation mechanisms and the effectiveness of two different methods operating under similar conditions. Effectiveness was evaluated by monitoring the change in aerosol mass concentration as a function of treatment time in 0.65 and 90 m3 chambers. Results show the effect of initial mass concentration and applied power density on coagulation rate.

EXPERIMENTAL PROCEDURES

1. Aerosol Generation for Turublent Agglomeration Tests

Turbulent agglomeration tests were conducted in a 90 m^3 sealed chamber, 4 meters in height. Sodium aerosol was generated by burning the metal in an electrically heated steel pot. The sodium pool, 260 cm^2 in area, produced a dense white aerosol that filled the chamber within 10 minutes after initiation of a fire. A one pound sodium pool fire would burn for 20-30 minutes. Decay in mass concentration with time was obtained with open-faced MSA 1106B high efficiency filter samples taken near mid-chamber height. Ten to 15 samples were taken over a 3 to 6 hour period at 18.4 lpm for 0.5 to 3 minutes. Less than 1% of the aerosol was removed by sampling.

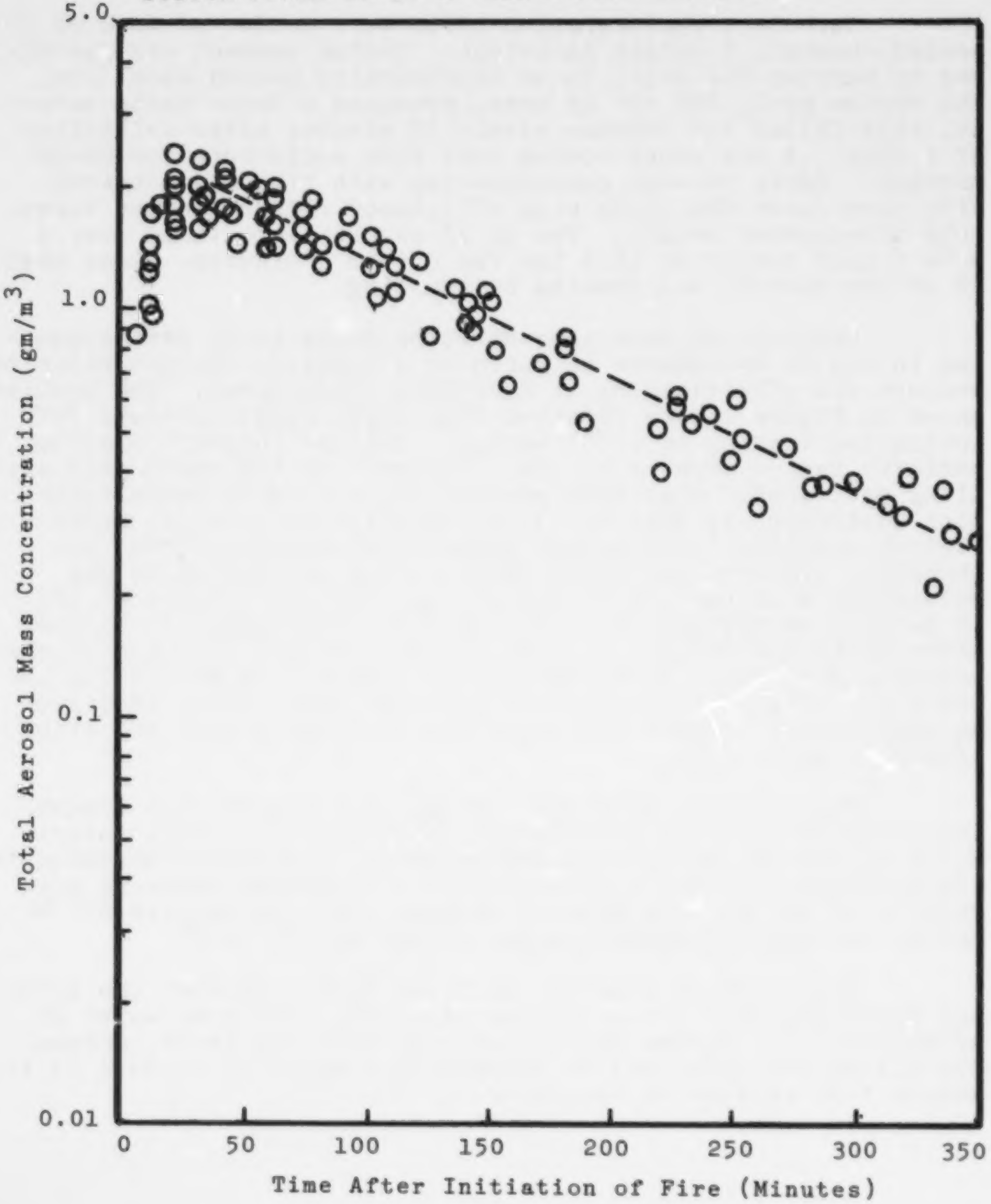
Undisturbed mass concentration decay tests were conducted in the 90 m^3 chamber to serve as a baseline against which to measure the effectiveness of turbulent enhancement. The results shown in Figure 1 were obtained from eight baseline tests following the burning of 1 lb. sodium. Initial chamber relative humidity ranged from 14 to 22%. Analysis of the unhindered settling data showed that this aerosol has the decay characteristics equivalent to that for 4 μm monodisperse particle under stirred settling. The actual mass median diameter (MMD) and geometric standard deviation (GSD) of the aerosol cloud was determined with the aid of eight-stage Andersen impactors and an aerosol centrifuge [1]. The size distributions, 20 minutes after initiation of a 1 lb. sodium fire for all of the generated aerosols were found to be nearly the same with a MMD of 2.1 μm and a GSD of 1.8. Measurements taken at later times indicated an equilibrium between sedimentation and growth with MMD stabilizing at about 3 μm.

Measurements were made of particle density and dynamic shape factor from the aerodynamic equivalent diameters determined by aerosol centrifuge and geometric parameters measured in the scanning electron microscopy [2]. Untreated aerosols aged from 34 to 376 minutes gave an average particle density of .78 gm/cm^3 and average dynamic shape factor of 1.37 [2].

Quantitative chemical analysis indicated that the aerosol particles were porous sodium carbonate with some water of hydration [3]. During the undisturbed settling tests, aerosol deposition per unit area of chamber wall was 0.10 to 0.01 of the amount that settled on the floor.

FIGURE 1

BASELINE CONCENTRATION DECAY PROFILE FOR 1.0 LB. SODIUM FIRES IN 90 M^3 AIR-FILLED CHAMBER



2. Turbulent Agglomeration Tests

A 7.5 HP, 3,000 cfm centrifugal blower (New York Blower Co., type GI, size 20) was placed in the center of the 90 m^3 chamber to induce turbulence. The blower was started shortly after peak aerosol concentration was reached. The time decay history of five such tests with peak aerosol concentration near 2 gm/m^3 is presented in Figure 2.

The blower was run with and without the scroll casing in place to distinguish the effect of centrifugal deposition from the effect of turbulence enhanced agglomeration and sedimentation. Operation with the scroll casing off takes away a major surface for deposition, though some cleaning credit must still be taken because of particle impaction on the fan blades. Removing the casing did not greatly change the decay profile from casing-on performance, as can be seen in Figure 2. The primary aerosol decay mechanism was, therefore, by enhanced agglomeration and sedimentation onto the chamber floor. Analysis of samples of wall, floor, and blade deposits indicated that aerosol deposition occurs, roughly, 28% on walls, 65% on floor, and 7% on the fan blades. Numerous large spherical particles ($d > 100$ μm) were found deposited on horizontal surfaces following each turbulence test.

Substantially faster decay rates were obtained with turbulent agglomeration when the peak aerosol mass concentration in the chamber was above 2 gm/m^3. One such tests in shown in Figure 3.

By using different pulley combinations on the blower, it was possible to generate different levels of turbulence in the chamber. Power densities, determined from measurements of fan speed and electricity consumption, are discussed in a later section. Assuming an electrical to mechanical energy conversion efficiency of 0.80, power densities in the range of 20-70 watts per cubic meters of chamber volume were utilized.

3. Acoustic Agglomeration Tests

Efforts to determine the effect of acoustic agglomeration on sodium combustion aerosols began with the search for a suitable and convenient source of high intensity sound. The first sound generating devices employed in this research were pulse jet engines [4]. A special pulse jet engine constructed by Ontario Research Foundation (Canada) was unable to provide a sound field of sufficient intensity for acoustic agglomeration experiments in the 90 m^3 chamber. These tests were described in Reference 3.

FIGURE 2

TURBULENT AGGLOMERATION AEROSOL CLEARANCE TESTS

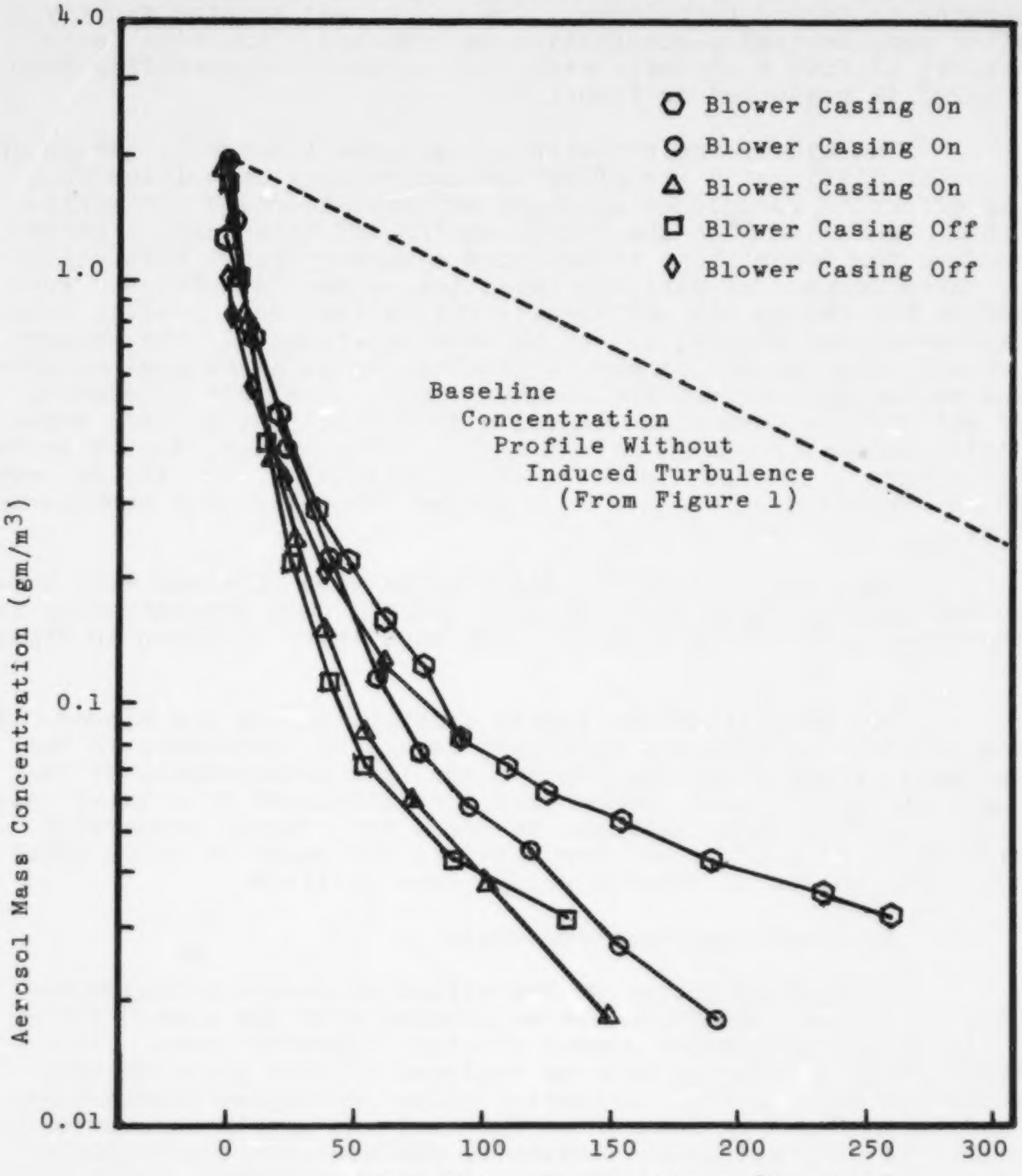
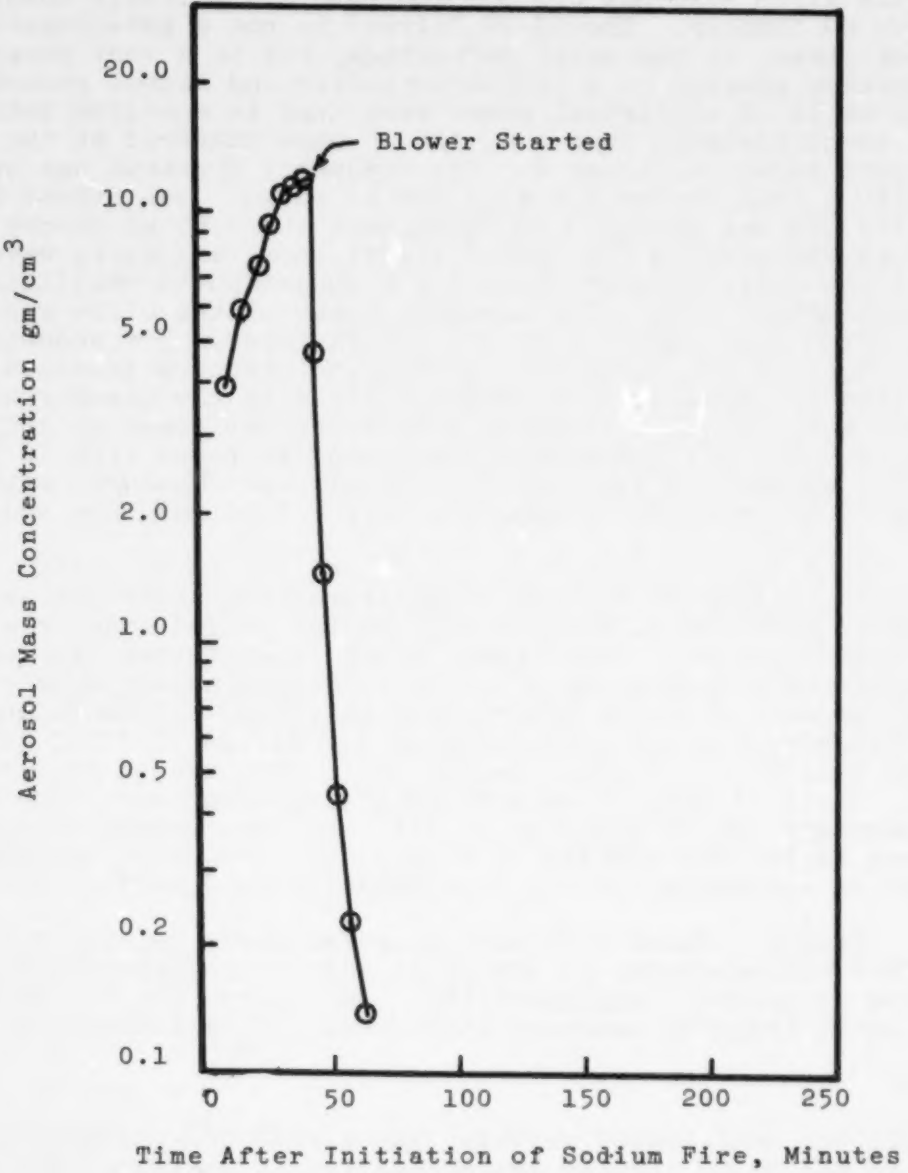


FIGURE 3

TURBULENT AGGLOMERATION FOR AEROSOL CONCENTRATION OF 12 GM/M^3



For later tests, a Federal CJ-24 electronic siren (a type employed in emergency vehicles) was used to generate intense sound fields in a steel chamber having a volume of 0.65 m^3 and an average internal height of 75 cm (Figure 4). A smaller chamber was needed because the acoustic power radiated by a single CJ-24 siren will not produce adequate sound levels thoughout a 90 m^3 chamber. The CJ-24 "siren" is not a gas-dynamic rotating siren, by the usual definition, but is a very powerful loud speaker powered by a PA-20A amplifier and signal generator. When 60 watts of electrical power were used to energize this siren, sound pressure levels of 145 dB were measured at the sampling port shown in Figure 4. The frequency spectrum was broad band with a peak in the 600 to 1,200 Hz range. Variations in intensity did not exceed 3 dB throughout the 0.65 m^3 chamber. The siren was used in the "yelp" signal mode, a complex wave form in the audible range which has a superimposed oscillation of approximately 5 Hz. The acoustic power of the siren was calculated from data supplied by the manufacturer. The acoustic power output of the siren for lower sound pressure levels was calculated by assuming a reverberant field in the chamber with constant wall losses. Acoustic tests were performed at 145 dB, 132 dB, and 125 dB, spanning a power density range from 10 watt/m^3 to 0.1 watt/m^3. Sound pressure levels were measured with a General Radio octave band analyzer in the flat response mode.

In a typical test, between 50 and 100 gm of sodium metal was heated rapidly to burning temperature in a steel pot and allowed to burn for 10 minutes filling the sealed chamber with a dense sodium aerosol. Open-faced glass fiber filter samples were obtained at mid-chamber height to determine the mass concentration over time for unperturbed and siren-on conditions. Each filter sample was restricted to 9.2 liters (1.4% of the chamber volume) to minimize depletion of the aerosol by sampling. Up to 11 samples were taken giving a maximum concentration decrease due to sampling of 14%. Because unperturbed and siren-on tests were sampled at roughly the same time intervals, changes in concentration due to sampling were identical for each.

Figure 5 shows mass concentration decay results for unperturbed sedimentation in the small chamber and sedimentation enhanced by acoustic agglomeration. The curves for the three sonic tests indicate moderate enhancement of sedimentation.

RESULTS

Data for undistrubed settling and turbulent agglomeration tests in the 90 m^3 chamber are summarized in Table 1. The basic parameters which were extracted from the concentration decay graphs for each test are initial aerosol mass concentration

FIGURE 4

0.65 M^3 CHAMBER FOR ACOUSTIC AGGLOMERATION EXPERIMENTS

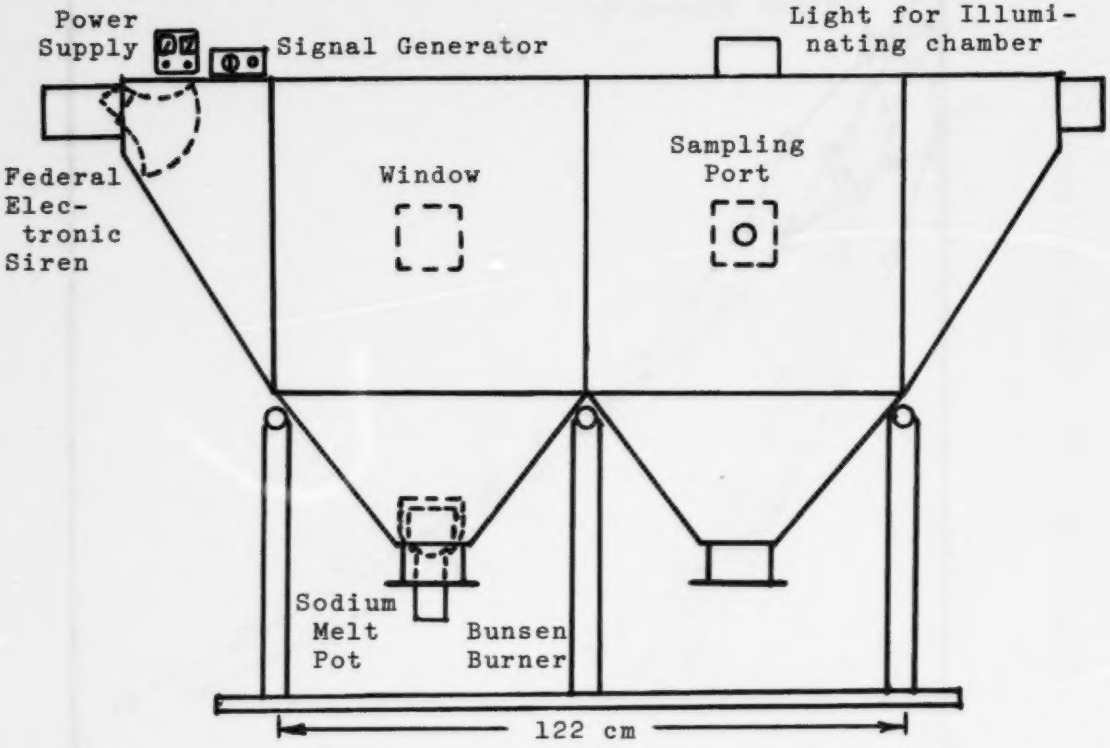

Power Supply
Signal Generator
Light for Illuminating chamber
Federal Electronic Siren
Window
Sampling Port
Sodium Melt Pot
Bunsen Burner
122 cm

FIGURE 5

ACOUSTIC AGGLOMERATION AND UNDISTURBED SETTLING IN A 0.65 M^3 CHAMBER

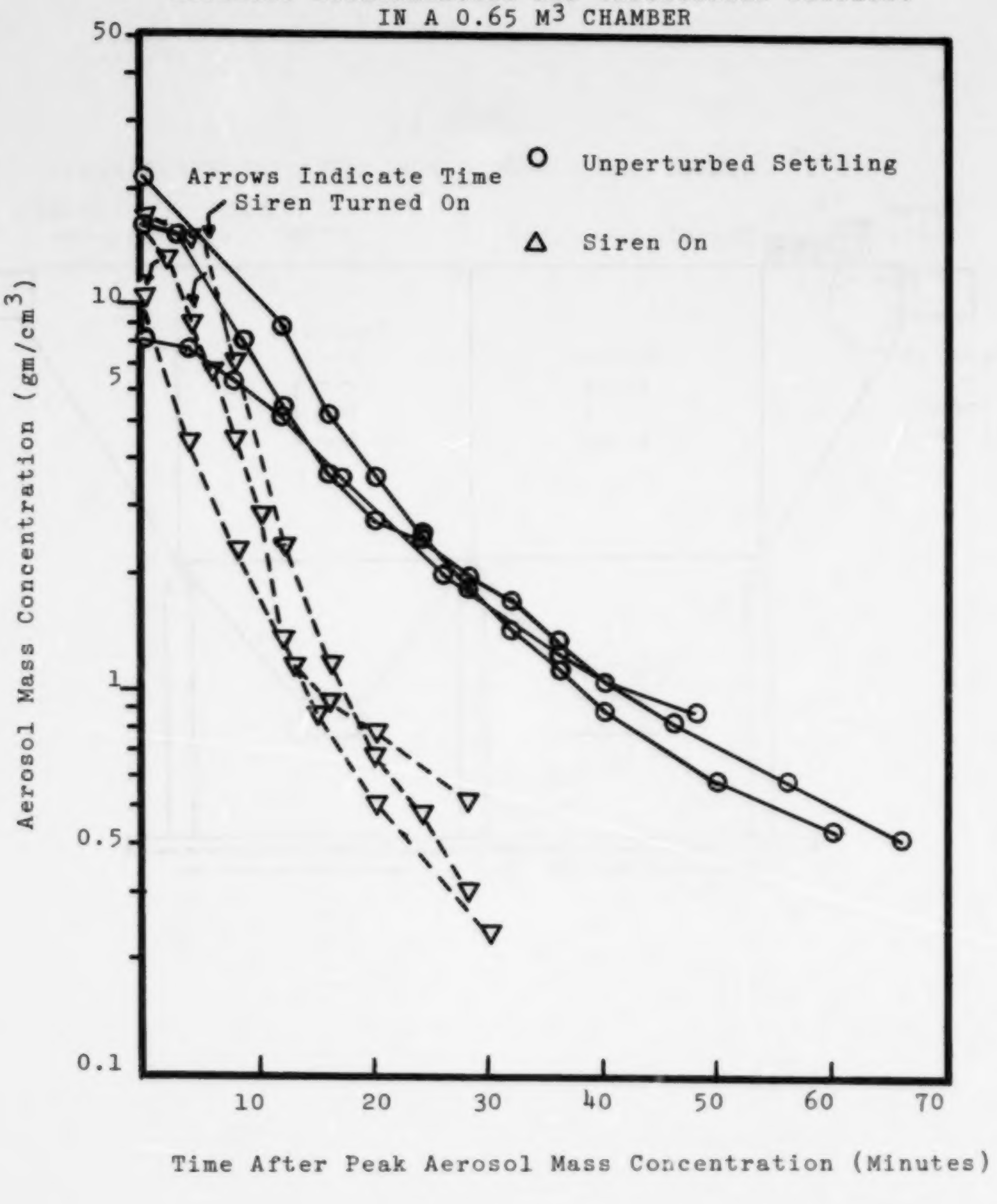


TABLE 1 - SUMMARY OF UNDISTURBED SETTLING AND TURBULENT AGGLOMERATION TESTS IN 90 M^3 VESSEL

Initial Aerosol Mass Conc'n (gm/m^3)	Time for $C/C_o=e^{-1}$ (min.)	Mass Fraction Removed in 5 min.	Monodisperse Equivalent Size, d* (μm)	Agglomeration Index I_A	Power Density watts/m^3
Undisturbed Settling Tests					
6.5	47	0.10	9.2	4.4	0
3.8	105	0.047	6.2	3.0	0
2.4	162	0.031	5.0	2.4	0
1.6	175	0.028	4.8	2.3	0
1.1	290	0.017	3.7	1.8	0
0.084	592	0.008	2.6	1.2	0
Turbulent Agglomeration Tests					
3.4	3.4	0.76	33.8	16.1	27
3.1	2.1	0.91	43.7	20.8	27
2.0	7.5	0.49	23.1	11.0	21
2.0	2.3	0.89	41.7	19.9	27
1.9	3.8	0.74	32.5	15.5	71
1.7	16	0.27	15.8	7.5	18
1.6	15	0.29	16.3	7.8	18
1.3	12.5	0.33	17.9	8.5	27
1.2	25	0.18	12.7	6.0	18
0.94	13	0.32	17.6	8.4	18
0.85	25.5	0.18	12.5	6.0	27

(C_o) and time (t_e) for the mass concentration to reach (C_o/e). For undisturbed settling, the initial mass concentration was considered to be the peak concentration reached. For the turbulence tests, initial concentration was defined as the measured concentration at the time turbulence was initiated, usually within a few minutes of peak chamber concentration. Data from undisturbed settling and acoustic agglomeration tests in the 0.65 m^3 chamber are similarly presented in Table 2 and defined in like manner.

Most mass concentration decay profiles gave straight lines when plotted semi-logarithmically, indicating exponential decay functions. Stirred settling conditions occur in the chambers due to thermal convective motion and energy from the agglomeration mechanisms themselves. It is well known that stirred settling of monodisperse aerosols produces exponential mass concentration decay. Even though the sodium aerosol is not monodisperse, it is extremely useful to approximate aerosol settling characteristics by the exponential decay of a monodisperse aerosol with 1/e decay time, t_e, measured directly from the concentration data. The mass fraction removed, f(t), in any standard time interval (e.g., 5 minutes, as in Tables 1 and 2) is readily determined from the relation,

$$f(t) = 1-e^{-t/t_e} \tag{1}$$

More significantly, the particle diameter of an "aerodynamically equivalent monodisperse aerosol" with the same settling characteristics, d_*, can be derived from the decay profiles. We have, for a stirred settling monidisperse aerosol of diameter, d_*, in a chamber of height, h,

$$C(t)/C_o = \exp\left(\frac{\rho {d_*}^2 g t}{18 \eta h \chi}\right) \tag{2}$$

where g = gravitation acceleration
η = gas viscosity
ρ = particle density
χ = aerodynamic shape factor

Equating the bracketed quantity in Equation 2 with t/t_e gives d_* as,

$$d_* = \left(\frac{18 \eta h \chi}{\rho g t_e}\right)^{1/2} \tag{3}$$

TABLE 2 - SUMMARY OF UNDISTURBED SETTLING AND ACOUSTIC AGGLOMERATION TESTS IN 0.65 M^3 VESSEL

Initial Aerosol Mass Conc'n (gm/m^3)	Time for $C/C_o=e^{-1}$ (min.)	Mass Fraction Removed in 5 min.	Mono-disperse Equivalent Size, d* (μm)	Agglomeration Index I_A	Power Density $watts/m^3$
Undisturbed Settling Tests					
21.5	13	0.32	7.6	3.6	0
16.5	11	0.37	8.3	4.0	0
8.2	20	0.22	6.1	2.9	0
6.8	23	0.20	5.7	2.7	0
Acoustic Agglomeration Tests					
14.2	3.3	0.78	15.1	7.2	9.1
10.5	4.9	0.64	12.4	5.9	9.1
9.0	5.3	0.61	11.9	5.7	9.1
7.6	7.0	0.51	10.4	5.0	0.46
4.2	28.5	0.16	5.1	2.4	0.091

Values for d_* computed for each test are given in Tables 1 and 2.

A useful parameter for characterizing both undisturbed and enhanced agglomeration tests is the "agglomeration index", I_A, defined here as the ratio of d_* to the initial aerosol MMD. Eight-stage Andersen impactor data indicate that within the first 20 minutes after initiation of a sodium pool fire, the MMD of the aerosol is about 2.1 μm. Using this reference MMD to form the agglomeration index, we have plotted I_A versus initial mass concentration for undisturbed settling tests and for turbulent and acoustic agglomeration tests as shown in Figures 6 and 7. No distinction is made in these figures between tests at different power densities. Because of the limited range of our power densities, strong statements about their relation to I_A are not justified.

Figures 6 and 7 also present trend lines fitted to the data. In each case, the trend line is constrained to pass through I_A= 1.0 at zero mass concentration, a constraint which has an obvious and strong physical interpretation. As expected from basic coagulation theory [5] each of the four trend lines shows increasing agglomeration indices with increasing mass concentration. Enhanced agglomeration rate leads naturally to an increased average particle size due to agglomeration in a given time period. By adding turbulent or acoustic energy to the aerosol, relative motion between particles is enhanced and agglomeration speeded up. For the range of power densities used, turbulence yields a higher agglomeration index than acoustic for the same initial mass concentration.

For the turbulence experiments and the acoustic tests, the following results of multiple regressions of agglomeration index with respect to initial mass concentration and power denisty were obtained. The point $I_A=1$ at $C_o=0$ was included in each case:

$I_A = 0.302 + 4.85\ C_o + 0.093\ P$ turbulent agglomeration

$I_A = 1.05 + 0.441\ C_o + 0.029\ P$ acoustic agglomeration

where C_o = initial mass concentration in gm/m^3
P = power density in watts/m^3

Significance tests of the regression equation coefficients showed that both concentration coefficients were significant ($P<0.02$) and both power density coefficients were not significant. It is apparent that turbulent agglomeration is more effective than acoustic agglomeration for a given initial mass concentration and that turbulent agglomeration is enhanced to a greater extent

FIGURE 6

AGGLOMERATION INDEX AS A FUNCTION OF MASS CONCENTRATION FOR TURBULENT AGGLOMERATION AND UNDISTURBED SETTLING TESTS IN A 90 M^3 VESSEL

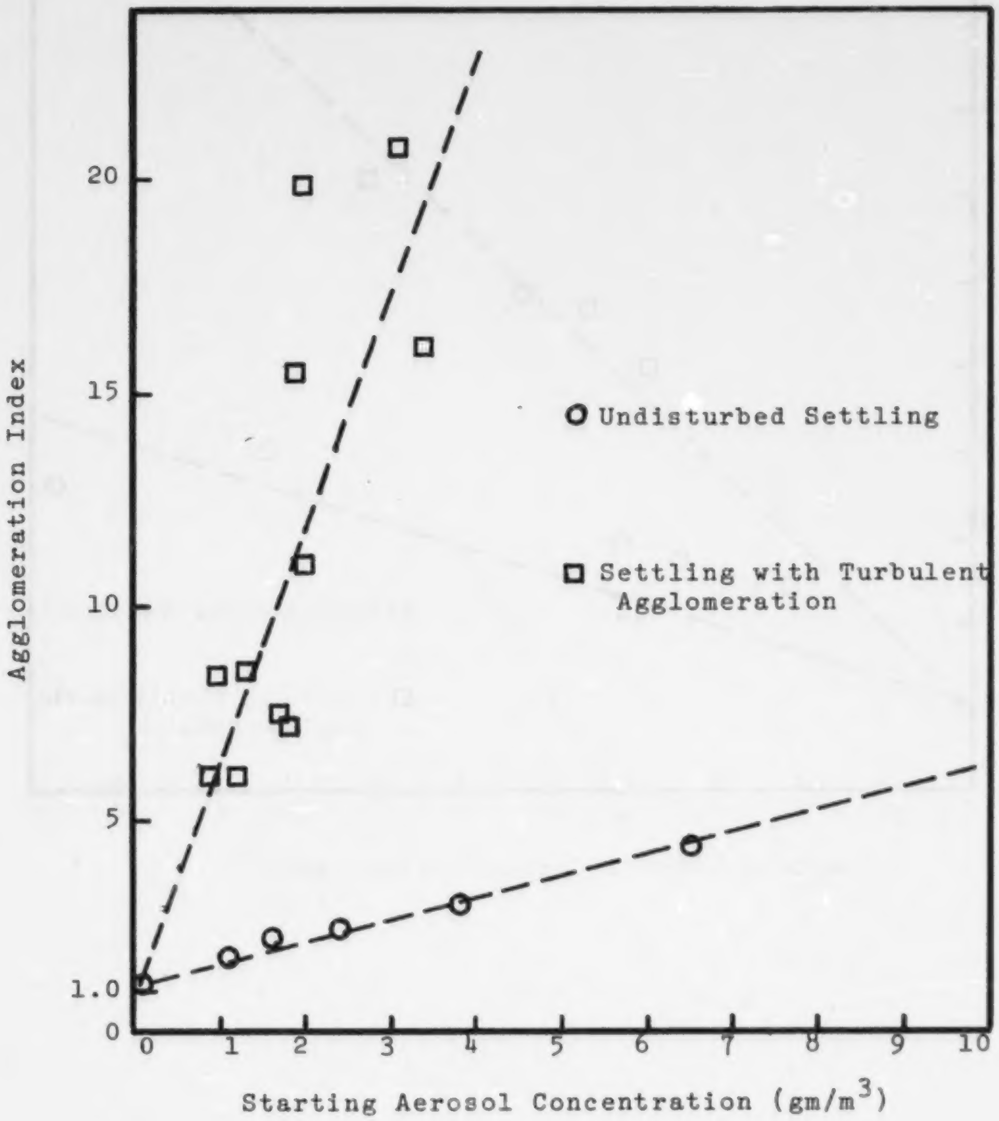
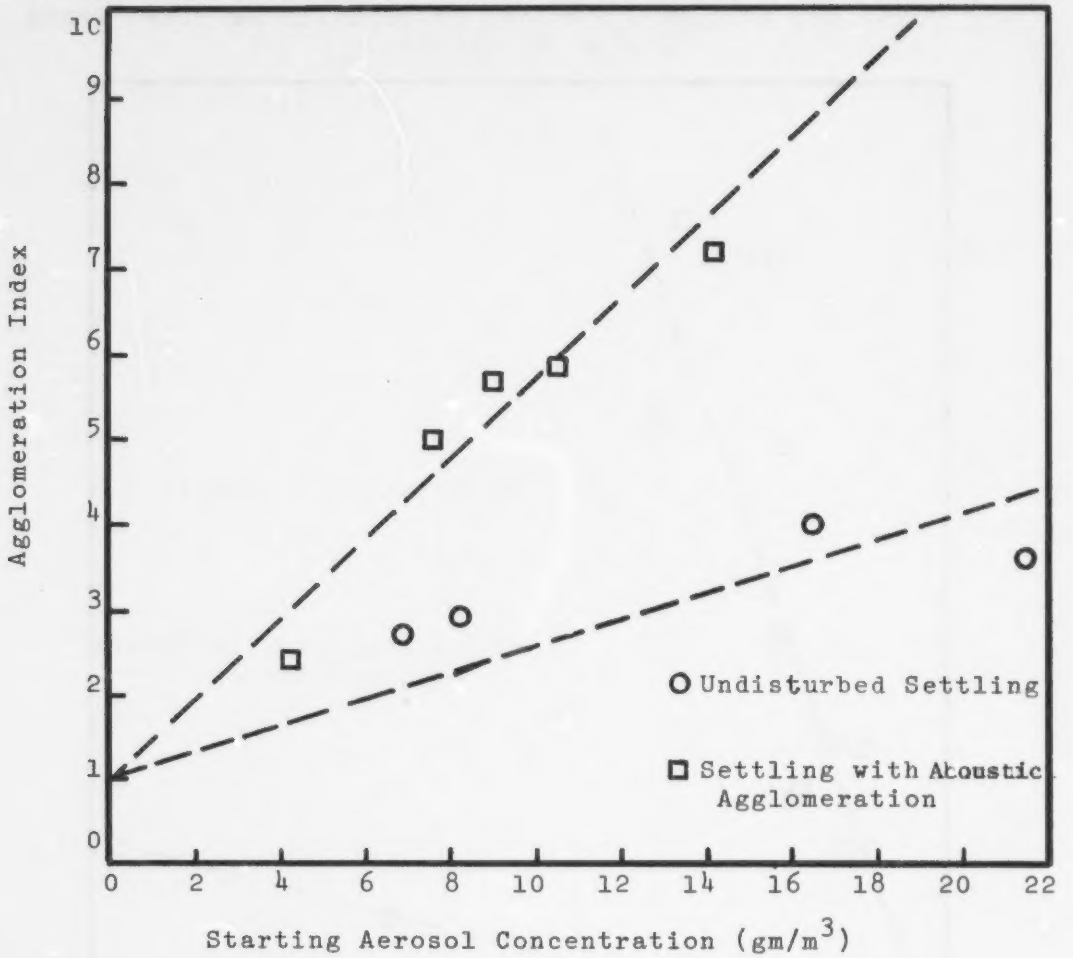


FIGURE 7

AGGLOMERATION INDEX AS A FUNCTION OF MASS CONCNETRATION FOR ACOUSTIC AGGLOMERATION AND UNDISTURBED SETTLING TESTS IN 0.65 M^3 VESSEL



for a fixed fractional increase in initial mass concentration than is acoustic agglomeration.

DISCUSSION AND CONCLUSION

The range of agglomeration indices found in our acoustic experiments is consistent with analogous measurements made by other researchers. Although no previous work compares directly with our research because sound sources were usually employed only for brief periods (seconds) on flowing rather than fixed volume systems, Scott [4] and Volk and Moroz [5], obtained agglomeration indices from 3 to 9 for sound pressure levels from 120 dB to 165 dB. Their mass concentrations were in the range 0.5 to 2.5 gm/m^3 and sound was continued from 2.5 to 40 seconds.

By using the concept of "scale of turbulence" noted by Fuchs [6], it is possible to gain some insight into why turbulent agglomeration does not show significant increases in effectiveness for significant increases in power density. The internal scale of turbulence, λ_o, used by Levich and referred to by Fuchs, is defined for a turbulent medium by,

$$\lambda_o = \left(\frac{\nu^3}{\varepsilon}\right)^{1/4} \tag{4}$$

where ν is the kinematic viscosity of the medium and ε is the power density per unit mass of the medium. For the range of power densities employed here the scale of turbulence, λ_o, ranges from 90 μm to 120 μm; a range unlikely to be associated with large changes in agglomeration index.

Turbulent motion and high intensity reverberant acoustic fields induce rapid agglomeration of sodium aerosols. The higher the initial aerosol mass concentration, the more effective are these agglomeration mechanisms which produce enhanced sedimentation. For undisturbed settling, sedimentation is similarly enhanced by increasing initial mass concentration. For the same energy expenditure turbulence appears to be a more effective agglomerating mechanism than acoustic energy at a fixed mass concentration. Increasing the applied power density of turbulence or sound produces smaller increases in agglomeration index than does a similar fractional increase in initial mass concentration. Both mechanisms show promise for direct application air cleaning systems. Because of the relative ease of generating turbulence versus generating a large-scale high intensity sound field, turbulent agglomeration may be preferred.

ACKNOWLEDGEMENT

Support of this research was provided by the U.S. Energy Research and Development Administration under Contract No. E(11-1)-2801.

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WRITTEN DISCUSSION

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Aerosol behavior within Liquid Metal Fast Breeder Reactor (LMFBR) containments is of critical importance since most of the radioactive species are expected to be associated with particulate forms, and the mass of radiologically significant material leaked to the ambient atmosphere is directly related to the aerosol concentration airborne within the containment. Mathematical models describing the behavior of aerosols in closed environments, besides providing a direct means of assessing the importance of specific assumptions regarding accident sequences, also serve as the basic tool with which to predict the consequences of various postulated accident stiuations. Consequently, considerable efforts have been recently directed toward the development of accurate and physically realistic theoretical aerosol behavior models.

These models have accounted for various mechanisms affecting agglomeration rates of airborne particles matter as well as particle removal rates from closed systems. In all cases, spatial variations within containments have been neglected and a well-mixed control volume has been assumed.

The most recent of these, the HAARM-2 code developed at Battelle's Columbus Laboratories[1*], is an extension of the HAA-3B code developed at Atomics International[2]. The HAARM-2 model differs from previous models in that it allows temporal variation of containment gas temperature, pressure, and temperature gradient normal to the containment walls. Also, settling velocities which are dependent upon the morphological properties of individual agglomerates are corrected using an experimental dynamic shape factor[3]. In addition, wall deposition by thermophoresis is included as an aerosol deposition mechanism and a calculated particle-particle collision efficiency is employed. Though not included previously in any model, terms describing agglomeration as caused by gas turbulence have been added to the HAARM-2 code so that calculations in support of this discussion could be made.

GENERAL THEORY

The theory used to develop the HAARM-2 code is based on solution of the governing integro-differential equation which describes the rate of change of particle concentration due to various agglomeration and removal mechanisms. This equation may be written in the following form:

$$\frac{\partial}{\partial t} n(x,t) = K_o \Big[1/2 \int_o^x \phi(\xi,x-\xi)\, n(\xi,t)\, n(x-\xi,t)\, d\xi$$

$$-n(x,t) \int_o^\infty \phi(x,\xi)\, n(\xi,t)\, d\xi\Big]$$

$$-n(x,t)\, R(x) + S(x,t) \quad , \qquad (1)$$

where

$\phi(x,\xi)$ = the normalized collision kernel predicting the probability of collision between two particles of volume x and ξ due to Brownian motion, gravitational settling, and, if included, turbulent gas motion

$K_o = 4kT/3\eta$ = the agglomeration rate constant

k = the Boltzmann constant

T = the gas temperature

η = the gas viscosity

---

[*] Numbers refer to references at the end of this discussion.

$x=\frac{4}{3}\pi r^3$ = volume of particle with radius r

$\xi=\frac{4}{3}\pi r'^3$ = volume of particle with radius r'

$n(x,t)$ = the size distribution function

$R(x)$ = the removal rate of particles produced by gravitational settling to the floor, diffusion to the walls (wall plating), and leakage

$S(x,t)$ = represents the source rate of particles input to the vessel.

The first integral in Equation (1) represents the formation rate of particles between the sizes x and x+dx as a result of collisions between particles of volumes ξ and $x-\xi$. Similarly, the second integral represents the disappearance rate of particles in the size range between x and x+dx due to collisions with all other particles.

The functional form of the collision kernel $\phi(x,\xi)$ depends upon the coagulation mechanisms present in a given system. In an enclosed containment vessel, possible mechanisms causing relative motion between particles, and thus coagulation, include Brownian motion of the particles, gravitational settling, and turbulent gas motion. In most analyses where more than one of these mechanisms is present, they are assumed to be separable and additive such that

$$\phi(x,\xi) = B(x,\xi) + G(x,\xi) + T(x,\xi) \tag{2}$$

the agglomeration and removal terms will be described in more detail.

AGGLOMERATION TERMS

Brownian Agglomeration

Current aerosol models use the Brownian collision parameter B in a form which can be written as

$$K_o\ B(x,\xi) = \frac{K_o}{2}(r + r') \cdot \left[\frac{r + a + be^{-\lambda r}}{r^2} + \frac{r' + a + be^{-\lambda r'}}{r'^2}\right] \tag{3}$$

where a, b, and λ are constants defining the Cunningham correction factor which accounts for the low Knudsen number effects present for small particles.

Gravitational Agglomeration

The collision parameter for gravitational agglomeration is based on consideration of the relative sedimentation rates for different sized particles and is dependent on the collection of small particles from the volume swept out by a large, rapidly settling particle. The gravitational agglomeration parameter is given by

$$K_o\, G(x,\xi) = \frac{2\pi\rho_p g}{9\eta}\, \varepsilon\,(x,\xi) \cdot \{|\, r\alpha(r) - r'\alpha(r')| \cdot (r + r')^3$$

$$+ a(r + r')^2] + b(r + r')^2 \cdot |\, re^{-\lambda r}\alpha(r) - r'\, e^{-\lambda r'}\alpha(r')|\} \qquad (4)$$

where

$\varepsilon(x,\xi)$ = the particle-particle collision efficiency

$\alpha(r)$ = the particle shape factor

ρ_p = the density of the particle material.

Turbulent Agglomeration

Particles are caused to collide in a turbulent gas by being suspended in a shear flow and by inertial effects cuasing relative motion between particles[4]. The shear and inertial components of the turbulent collision parameter are termed T_1 and T_2, respectively. These should appear as $T = (T_1^2 + T_2^2)^{1/2}$, but have been assumed additive as $T = T_1 + T_2$ to permit analytical integration. The expression for the shear induced component of the turbulent collision parameter is

$$K_o T_1(x,\xi) = \varepsilon(x,\xi)\left[\frac{8\pi\varepsilon_T\rho}{15\eta}\right]^{1/2} (r + r')^3, \qquad (5)$$

and the inertial component of the turbulent collision parameter is

$$K_o T_2(x,\xi) = \varepsilon(x,\xi)\, \frac{4\pi_p\sqrt{2\pi}}{9\eta}\left[\frac{1.69\varepsilon_T{}^3\rho}{15\eta}\right]^{1/4}$$

$$|(r)\,(r^2 + ar + bre^{-\lambda r}) - \alpha\,(r')\,(r'^2 + ar' + br'e^{-\lambda r'})|$$

$$\cdot\,(r + r')^2, \qquad (6)$$

where ε_T is the rate of turbulent energy dissipation per unit mass (cm^2/sec^3), and ρ is the gas density.

Removal Terms

In Equation (1), R(x) represents the removal terms. HAAM-2 accounts for any of the four possible mechanisms: (1) removal at the vessel floor due to gravitational settling, (2)Brownian diffusion to the walls (wall plating), (3) thermophoretic deposition on the walls, (4) leakage out of the vessel. Mathematically formulated

$$r(x) = G_R\, r(r + a + be^{-\lambda r}) + P_R\, \frac{1}{r^2}\,(r + a + be^{-\lambda r}) + R_L + T_R \qquad (7)$$

where

$$G_R = \frac{2g\alpha(r)\rho_p}{9h\eta}$$, the gravitational settling constant

$$P_R = \frac{kTA_w}{6\pi\eta V\Delta}$$, the plating coefficient

R_L = leak rate from the vessel

$$T_R = \frac{A_w}{V}U_t$$, thermophoretic deposition coefficient.

further, h is the vessel height, A_w the wall area, Δ the distance from the wall over which a particle concentration gradient exists, U_t the thermophoretic velocity, and V the vessel volume. In the HAARM-2 code the assumption has been made that the final term in the slip correction factor is negligible and hence for both gravitational settling and agglomeration the constant is taken as zero.

The method of moments was used to solve the governing integro-differential equation, Equation (1), and the solution technique has been described previously[1,2,5]. Basic to this method is the assumption that the particle size distribution may be represented by a log-normal distribution at all times and the parameters describing the particle population (number concentration, mean size, and geometric standard deviation) are derived at the end of selected time intervals in terms of the moments.

RESULTS

A series of calculations were performed to compare the HAARM-2 aerosol behavior model with the experimental measurements of Hinds, Mallove, and First. Of particular interest was the comparison of the predictions and experimental results for the case of turbulent induced augmentation of the agglomeration rate since this is the first such comparison ever possible. Of course, the base cases with no added turbulence are of importance in establishing a reference point for comparison with the turbulent cases.

The base cases with no added turbulence were calculated in three ways. First, the instantaneous release of sodium oxide aerosol was assumed to occur at the end of the sodium burning period of 20 minutes. The characteristics of the aerosol at this time were taken as aerodynamic mean diameter = 2.1μm, effective density = 0.78 g/cm^3, shape factor = 1.37 and mass concentration = 2.0 g/m^3. The HAARM-2 code employs mass equivalent particle diameters and makes its own corrections to particle mobility accounting for effective density and shape factor combined. Therefore, an initial mass equivalent radius, $\bar{r}$, of 1.4 μm was used (calculated by correcting the impactor determined aerodynamic diameter for density and shape factor), a geometric standard deviation, σ_g, of 1.8 was assumed for the distribution, the material density, ρ_p, was taken as that for hydrated Ha_2CO_3 or 1.5 g/cm^3, and a constant correction for density, α, of 0.33 was assumed.[3] These values are consistent with those reported by Hinds, Mallove and First for the particles at the end of the burning period.

A second similar case also assumed an instantaneous source and identical conditions except that the value used for the aerodynamic mean radius was 1.5 μm as reported by Hinds, Mallove and First for times after the end of the fire which corrected for effective density and shape gives the value of $\bar{r}$ =2.1.

A third case followed the procedure usually used to predict aerosol behavior following a sodium fire. A constant particle source rate was assumed over the 20-minute burning period. In this case the source particles were assumed to have a mass mean radius of 0.5 μm, a material density of 2.27 g/cm^3, and a geometric standard deviation of 2. The effective density for the particles was calculated within the program as a function of the number of particles forming an agglomerate.

The results of HAARM-2 code calculations for these three cases are shown in Figure 1. It is seen that the predicted airborne mass concentrations are consistently higher than the experimental values. This is consistent with the nature of the HAARM-2 code which was intended to provide conservative estimates for nuclear reactor safety analyses and, therefore, employs procedures which always lead to slight overestimates of airborne concentration. Nevertheless, the aerosol behavior model shows reasonable agreement with these most useful data.

To compare the HAARM-2 code calculations with the experimental results when the blower was operated, the terms for turbulent agglomeration were added to the HAARM-2 code. To simplify the integration for this comparison it was assumed that the particle-particle collision efficiency, ε, which appears in Equations (5) and (6) was a constant rather than a function of particle size. This constant is taken as unity in the turbulent agglomeration terms but is included as a function of particle size in the gravitational agglomeration term. The calculations to show turbulent agglomeration effects were based on a value for ε_T, the turbulent energy dissipation rate, equivalent to 45 $watts/m^3$ which is the mid-point in the range of 20-70 $watts/m^3$ reported by Hinds, Mallove, and First for their experiments. Other assumptions were the same as for the baseline case calculations using $\bar{r}$ = 2.1 μm.

The results of the calculation for turbulent agglomeration are shown in Figure 2. It is seen that the prediction is on the low wide but adjacent to the range of experimental results. This is reasonable agreement but will probably be improved when the HAARM-2 code is modified to include a size dependent collision efficiency, ε, in the turbulent agglomeration terms and some accounting is made for overestimates resulting from the additive procedure which was used for agglomeration rates in the form $T=T_1+T_2$. The agreement between theory and experiment is very encouraging and shows promise of future improvement.

CONCLUSIONS

The data reported by Hinds, Mallove, and First are very useful as additional confirmation of sodium aerosol behavior in enclosed vessels and in

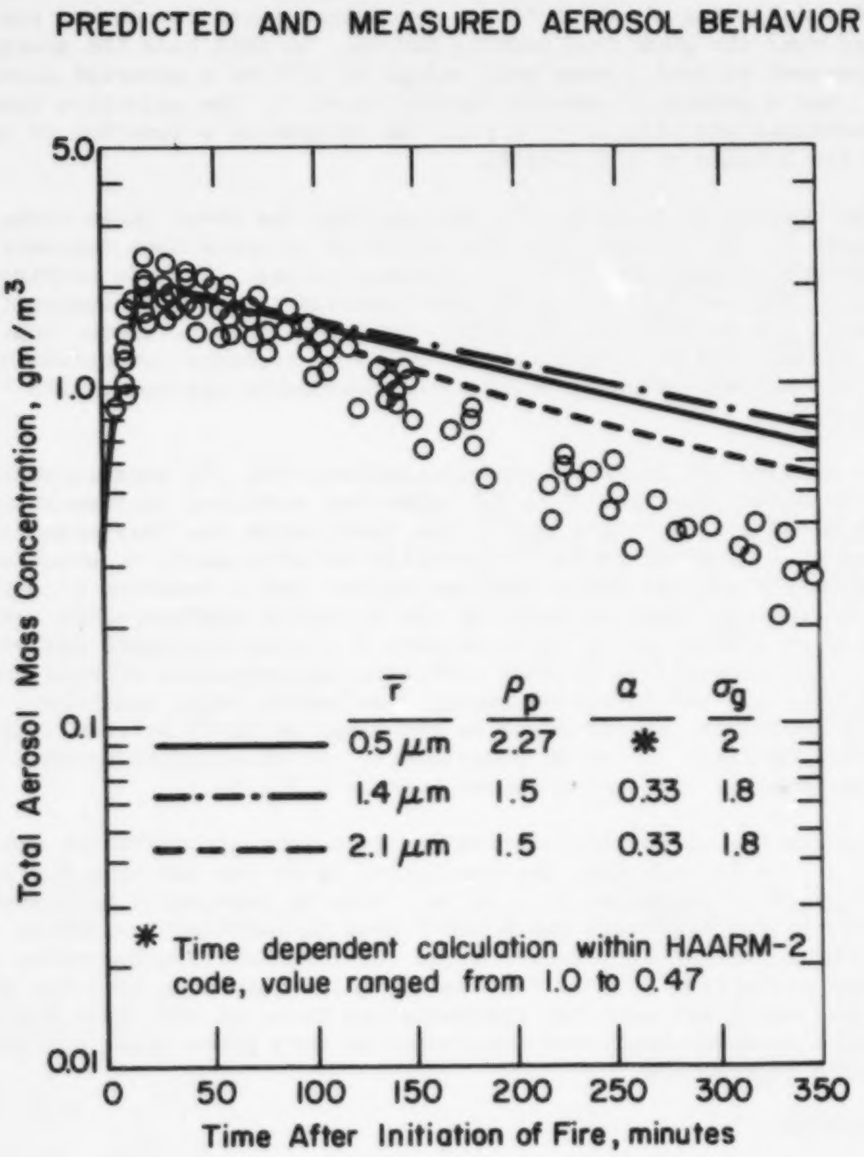

FIGURE 1. PREDICTED AND MEASURED AEROSOL BEHAVIOR

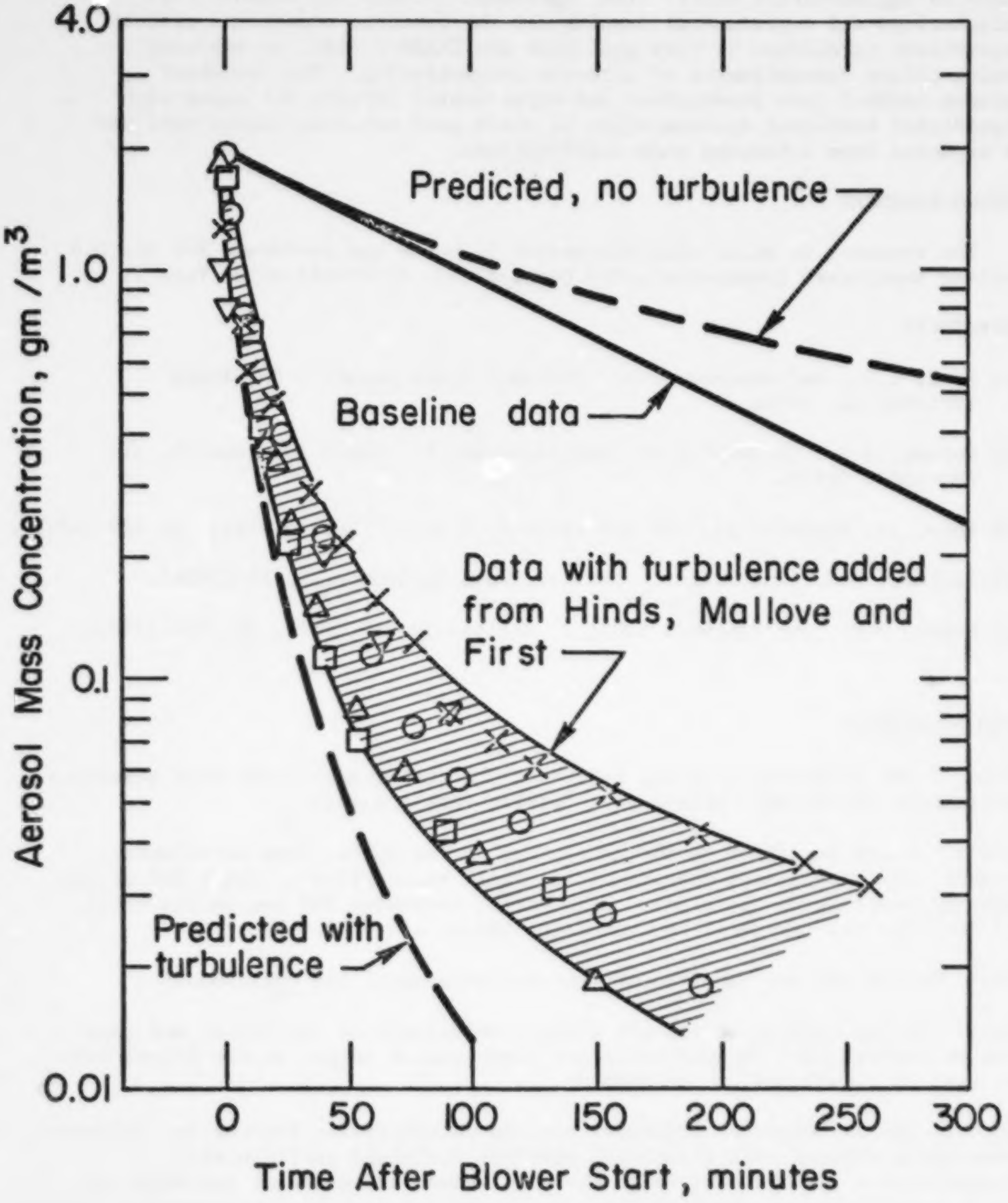

FIGURE 2. COMPARISON OF PREDICTED AND MEASURED TURBULENT AGGLOMERATION EFFECTS

particular provide a most useful basis for evaluating the effects of turbulence on agglomeration rates. The agreement between the HAARM-2 code calculations and experimental results for the baseline conditions without significant turbulence is very good with the HAARM-2 code, as expected, giving slight overestimates of airborne concentration. The agreement between HAARM-2 code predictions and experimental results for cases with significant turbulent agglomeration is quite good but some improvement can be expected from scheduled code modifications.

ACKNOWLEDGEMENT

The research on which this discussion is based was performed for the U.S. Nuclear Regulatory Commission under Contract No. AT(49-24)-0293, Task 7.

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OPEN DISCUSSION

Shaw: I was wondering, Jim and Bill, how the turbulence - both with acoustics and without acoustics - affect your cutting into the wall.

Hinds: We saw an effect on the deposition on the walls from turbulence, however, the predominant deposition was still on the floor. About 65% of the material ended up on the floor. Half of the remaining 35% was on the walls, and the rest was trapped directed on the blades of the device.

Shaw: Do you see any significant increase because of the turbulence?

Hinds: We had between 1% and 10% without turbulence on the walls, and then went up several fold because particles large enough to get in are driven onto the wall by turbulence on the chamber.

Cooper: The different coefficients for the concentration for the two different cases might suggest that there were somewhat different particle size distributions. So that, although the work shows the turbulent agglomeration can be significant, perhaps you have to be careful in making a comparison between the turbulent agglomeration in the big room and the acoustic agglomeration of the smaller chamber. Especially when these concentration coefficients are so different.

Hinds: We measured the size distribution in both cases. In the case of the small chamber, the burn took place very rapidly, about 2 minutes versus 15 minutes in the other one.

We waited until the fire ended before we made any measurements or treatment. At that time we had roughly the same size distribution. But it is a younger aerosol; the aerodynamic sizes were roughly the same.

Benarie: How did you monitor or control humidity in the chamber? Because you are working with a very humidity - sensitive aerosol, the conditions of a few percent change in humidity can cause a sizeable change in turbulence. Your experiments end up to what I am going to say on Friday morning.

In fiber filtration, we generally do not consider that air in the gates entering the filter develops a turbulence or anything like a turbulence. I will show you how this occurs. The turbulence contributes a large measure of agglomeration and leads to a better performance of filtration because parts of the filter are already agglomerated. So in developing filtration mechanisms which are most formed in uniform and laminar flow, a turbulent agglomeration should also be considered.

Hinds: The effect of humidity is important to a hygroscopic aerosol. In the runs that I described earlier, the humidity ranged from 16% to 24%. We found that in that range, based on our baseline runs, we got very reproducible results. When we went to 35% to 50% humidity, these results were quite different; by a factor of two in mass concentration.

Scott: I am very interested in your results. I noted that the highest sound pressure levels were 145 dB and I am intrigued that you caught anything there. You obviously caught something, but you will find that there is a marked increase in agglomeration ratio up towards 155 dB.

I think it would be interesting if you would introduce some other runs. It would be interesting if they were done at the same time and in the same chamber. It would also be interesting if you had three conditions. 1) Pure turbulence 2) Pure acoustic 3) A combination of turbulent and acoustic. Our preliminary results show that the third condition is where things really start to happen.

My last comment refers to power. I think you said that "turbulent agglomeration, for the power requirements, would be satisfactory without the acoustics." But we hope that this would only be used in the event of a catastrophe. At that point power would not be a criteria at all and should have no relevance. Would you please comment on this?

Hinds: On power requirements, your statement is true. However, in the case of an emergency, power may not be available. But slight changes in efficiency are really irrelevant in this case.

On sound pressure levels, we were aware that these were on the low end of the scale and we weren't able to reach the higher pressures. But we felt that if we could get above 140 dB we would have something that was worthwhile. We did reduce it down to 125 and didn't observe a great deal of difference.

The combination of systems is worth note and we are looking into it.

Lamb: What is the feasibility of artifically increasing the aerosol concentration? I'm thinking of a case in an emergency where you would blow a harmless aerosol which would scrub.

Hinds: If we could artifically increase the concentration we would be much better off. This is worth considering given the shape of those curves. In terms of the particular situation with the reactor, we have many thousands of pounds of non-radioactive sodium in the thing. All you have to do is open a valve, let sodium burn and you have got a lot higher concentration. But if you have an accident with a lot of sodium around, you just don't want to burn more. So this may not be very realistic.

Baril: Could you tell us a little bit about your sampling techniques. Where in the chamber did you take your samples?

Hinds: We sampled about mid-height in the chamber through a 1 1/2" diameter opening at about 20 liters/min. The sampling tube is 4 inches long and has a filter. We have a valve that we could shut off to isolate the chamber. We sampled directly under the filter. The samples were then weighed within a matter of seconds after removal to prevent any inclusion of moisture. The samplings were brief, lasting between one and three minutes, and spaced about 10 to 15 minutes apart. All were done at one height in the chamber.

Initially, when we burn the sodium, we see stratification. The hot sodium rises to the top, but there is enough convective energy from the burning sodium to mix it well. When the tests were taken, there was no stratification.

Oder: Will increased humidity enhance sedimentation, in which case, it is very easy to boil water without fear of loss of sodium?

Hinds: Yes. You could inject steam. However, if you have metallic sodium present this would be dangerous because of the evolution of hydrogen gas. So you have a tradeoff of blowing the thing up and getting the air back out.

Löffler: We sometimes use acoustics in liquids for agglomeration. I would like to know if there is any maximum or optimum power input or perhaps a final size of agglomerate which you can obtain if you put more energy into the system?

Hinds: The mechanisms involved in the scale of turbulence is such that it would tend to scale surfaces on the order of about 100 microns. It would also tend to break up particles that were long on that order of dimension and you would end up with a kind of spherical shape.

As for turbulence in the break-up of agglomerates in acoustics, I think David Scott would be more qualified to respond.

Scott: If you do break up particles in an aerosol, it is clear that it is the big ones that you are breaking up. It also makes sense, almost intuitively, that if you have got some long stringy thing that you are going to break up, it would break approximately in the middle, rather than chip little bits off the end. So it seems that you could bring the agglomerate to too large a size and break it into approximately equal size particles and you would kick up the number density which is desirable.

If you are talking about a solid particle in a liquid, that means you would be propagating through a nearly incompressible fluid. For a given acoustic field and intensity, the wavelengths are totally different. The amplitude, which is the critical thing, would be much smaller than in a gas. These things occur because you do have a dynamic motion. In a liquid, the dynamic motion would be much less than in a gas unless it was a bubbly liquid where you have sort of an analog thing too.

Löffler: A most important thing would be the stress which acts on the agglomerate and the forces, for example, adhesion forces. In the case of turbulent flow you will not get shear stress for particles in the range of one to five microns which will break the agglomerates. However, in acoustics it may be possible.

Hinds: I would tend to agree. The situation of mechanically generating turbulence gives a more intense turbulence than you would find in a normal turbulent flow.

Yang: Another note on the relation between the frequency of excitation and the size that you have. In turbulence in the case of a liquid in a tunnel you have an additional mechanism, a Reynolds stability. You can stretch the filaments and they will break up. It depends on the size and adhesion force.

NEW APPLICATIONS OF ACOUSTIC AGGLOMERATORS IN PARTICULATE EMISSION CONTROL

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INTRODUCTION

In order to meet the national goal of energy self-independence, expanded research and development on coal-fired and nuclear power generating plants have become a national priority. Numerous new programs have been established by various Federal agencies to seek answers to problems associated with how to generate electric power with low-grade(high sulfur) coal, to improve power plant efficiency, to increase our capacity in nuclear power generation, and above all to accomplish all these with minimum environmental degradation. These new programs lead to the new engineering problems such as the control or suppression of airborne dust particles in very unconventional situations - the combination of high temperature, pressure, or corrosive environment - in which the conventional air-pollution abatement equipment cannot function effectively. In many of these situations it appears that acoustic agglomerators possess certain characteristics which seem very attractive for dust-particle control in several of these new programs. These will be discussed first in the following sections, followed by the descriptions of some large sonic generators which may be suitable for these new applications.

Acoustic agglomeration is not a new idea. As early as 1931, Patterson and Cadwood [1] studied the effect of sound on aerosol dispersion. A great many papers appeared in the literature of the 1950's and these papers were summarized in the well-known book on sonic agglomeration by Mednikov, published in 1963 [2]. Since then however, there has been an almost complete vacuum in which little interest has been shown on the sonic agglomeration of aerosol particles. Shirokova reviewed acoustic agglomeration in 1973 with virtually no new experimental data. A series of papers were published in the USSR concerning the theoretical treatment of aerosol behavior in an acoustic field [3-10]. A group at the University of Ontario proposed the idea of using an acoustic field to enhance dust collection by filters [11]. At about the same time Scott and his associates worked on the idea of using a progressive wave for industrial aerosol emission control [12-15]. Their work shows that when properly designed, acoustic agglomerators can be economically competitive with more conventional devices for special appications in which conventional devices are not effective.

It is important to emphasize here that sonic agglomerators cannot compete in applications where the conventional devices already have been successfully used. Thus, in a recent report by the Midwest Research Institute [16], it is found that sonic agglomerators are too expensive for conventional pollution control because of the high power consumption in the generation of strong acoustic fields.

However, the sonic agglomerators have some characteristics which make them uniquely suitable for certain special applications. These include the acoustic conditioning of fine particulate emissions (particle diameter between 0.2-2μm) the suppression of radioactive sodium-fire aerosol in hypothetical Liquid Metal Fast Breeder Reactor (LMFBR) accidents, and the control of particulate emission under special conditions where conventional abatement techniques are not applicable. The objectives of this paper are to describe these new acoustic applications and to identify the predominant sonic agglomeration mechanisms for each application.

LMFBR SODIUM-FIRE AEROSOLS

In the development of the Liquid Metal Fast Breeder Reactor (LMFBR), a major safety problem concerns the suppression of the radioactive sodium-oxide aerosol particulates produced in a hypothetical core-disruptive accident. Assuming that a small fraction of the airborne sodium aerosol leaks into the environment, radiological analysis of the site boundary does show that inhalation of aerosol particles containing plutonium oxides may be a major concern. This is different from the problem associated with the Light Water Reactors (WR) in which the airborne thyroid doses from radioiodine is of principle concern. To minimize the leaking of these radioactive aerosols into the environment through small cracks, pipe fittings and so on, the effective suppression of these aerosol particles inside the containment building in case of such an accident is probably the highest priority in the development of LMFBR emergency air cleaning systems.

This so-called sodium-fire aerosol is produced from the combustion of liquid-metal sodium coolant in air. Sodium combustion can take place either rapdily through, for example, a ruptured pressurized pipe; or from a slow spill of sodium on to the reactor floor. In the former case it is called a spray fire, and in the later case a pool fire. A review covering both types of these fires can be found in the literature [17].

The sodium fire aerosol which consists primarily of sodium peroxides, have diameters predominantly in the range of 2-20μm. Owing to the constant thermal convective stirring, these relatively large particles stay airborne for a long period of time, say two or three days, with a very high total mass-loading (in the range of 1-50 g/m^3). In the recent report entitled, "Evaluation of Air Cleaning Concepts for Emergency Use in the LMFBR Plants", twenty-four different air cleaning systems were investigated and these are listed in Table 1. Among these, direct in-containment acoustic agglomeration is rated to be number one in terms of effectiveness, and number four in terms of fabrication, as shown in Table 2.

TABLE 1

System Number	Description	System Number	Description
SR-1	Prefilter-HEPA	SD-14	Liquid Sprays
SR-2	Deep Bed Graded Media	SD-15	Powder Discharge
SR-3	Sand Bed	SD-16	Foam Encapsulation
SR-4	Bag Filter	SD-17	Acoustic Agglomerator
SR-5	Cyclone	SD-18	Direct Electro. Precip.
SR-6	Cyclone plus HEPA	SC-19	Comb. Acoustic plus Cyclone
SR-7	Wet Scrubber	SC-20	Comb. Powder plus Filter
SR-8	Fluidized Bed		
SR-9	Acoustic plus Cyclone	CF-21A	Large HEPA, Charcoal HEPA
SR-10	Mechanical Separator	CF-21B	Small HEPA, Charcoal, HEPA
SR-11	Settling Chamber	CF-22	Wet Scrubber, HEPA, Char.,HEPA
SR-12	Electrostatic Precipator	CF-23	Elec.Precip.,HEPA, Char.,HEPA
SR-13	Steam Conditioning	CF-24	Sand Bed, HEPA, Char., HEPA

TABLE 2

RANKING OF TWENTY EACS CANDIDATES BY TOTAL SCORE AND BY CRITERIA GROUPS - SINGLE CONTAINMENT PLANT

Overall Rank	System	Total Score	Ranking by Criteria Group*					
			A	B	C	D	E	F
9	SD-17 In-containment Acoustic	309.5	1	7	15	16	16	4
14	SD-15 In-Containment Powder	296.4	2	16	11	18	14	6
11	Sc-19 In-Cont.Acoustic +Recirc.Cyc	304.6	3	8	16	15	11	9
13	SD-14 Liquid Spray	296.8	4	10	19	11	13	1
16	SC-20 Powder + Recfrc. HEPA	288.7	5	18	14	17	3	14

* A- Effectiveness, B- Reliability, C- Compatibility, D- Credibility, E- Flexibility, F- Fabrication

This suggests that if R & D effort can improve the credibility, reliability and compatibility of this concept,.acoustic agglomeration may be one of the top contendors for such emergency application.

CONTROL OF FINE-PARTICULATE EMISSIONS FOR STATIONARY SOURCES

The use of mass weight is an acceptable practice in particulate emission control since the current regulations are completely based on emission mass. Missing from the current air quality criteria is the particles sizes, which is probably the most important parameter in the discussion of environmental effects of particulate emissions. Large particles (diameter larger than 2μm) constituted the major mass fraction of most emissions sources. Although they represent a significant public welfare problem by contributing to the soiling and unsightly smoke plumes, these large particles are not a major health problem since inherently they tend to deposit in the upper airways of the human respiratory system.

It is the fine particles (diameter less than 2μm) that usually penetrate very deeply into the pulmonary airways and cause permanent health effects. Furthermore, these fine particles, when released into the atmosphere, stay airborne for an extended period; in contrast to the large particles, most of which are quickly brought back to the earth's surface by either direct fallout or by precipitation scavaging.

Large particles can be effectively controlled by conventional equipment such as cyclones, filters, scrubbers, electrostatic precipitators, etc. However, these devices have relatively poor efficiencies for the control of fine particles with diameters ranging from 0.2-2μm.

Since acoustic agglomeration has as its principle advantage the ability to control particles of any size without a minimum in the collection efficiency, it is potentially attractive for use as a preconditioner of the conventional emission control devices for the effective removal of fine particles. In other words, acoustic agglomerators offer distinct advantages in hybrid-unit applications in which the acoustic agglomerator is used upstream from the conventional devices. In this way acoustic agglomerators can be used to precondition the size distribution of the particles which permits a more effective emission control of fine particles. The purpose of the acoustic preconditioner is to shift the maximum of the dust particle size distribution to a larger particle size, leading to a more effective removal of the submicron particles by electrostatic precipitation.

It is important to emphasize that acoustic agglomeration has been used in the past for the control of dust particles, as summarized by a recent EPA report [16]. The traditional application of acoustic agglomeration for such applications is not considered to be economical because of the high power consumption required for the production of a strong acoustic field. Typically, an acoustic intensity of 160 db or higher is required. However, this must be distinguished from the application for fine particle control. In other words, acoustic agglomerators cannot compete with conventional devices for emission control on a weight basis. It should only be used for the effective control of fine particles with diameters between 0.2-2μm in which conventional devices are not effective. For such applications a relatively high frequency acoustic field must be used so that only fine particles are oscillating while particles with diameters larger than 2μm are relatively stationary. Thus, only the oscillation of fine particles consumes acoustic energy. Also the power consumption for the production of the acoustic field can be significantly reduced, and the use of acoustic agglomerators can be economically attractive.

EMISSION CONTROL OF A FLUIDIZED BED BOILER WITH A GAS TURBINE-STEAM TURBINE COMBINED CYCLE

To meet the requirement of producing adequate amounts of electric power with a minimum degradation of the environment, it has been increasingly important to find a way of burning the high-sulfur coal and at the same time, to remove SO_2 from the flue gas before release into the atmosphere. Among the concepts being developed, the idea of fluidized-bed combustion is rapdily gaining favor. In this case the high-sulfur coal is injected and burns in the bed at a temperature of 1450-1750°F. The SO_2 released during the burning of the coal is reacted with lime and in the presence of excess oxygen in the bed forms calcium sulphate. Fresh, crushed high-calcium limestone (dolomite limestone) is injected continuously into the bed and the sulphate lime is removed continuously.

Several concepts of the fluidized in-bed combustors have been considered for power plant applications [19-22]. Among these the ideas of using a pressurized, fluidized bed combustor in a combined gas-turbine and steam-turbine cycle is particularly attractive, as demonstrated in Fig. 1. Steam for operating a turbine is raised by

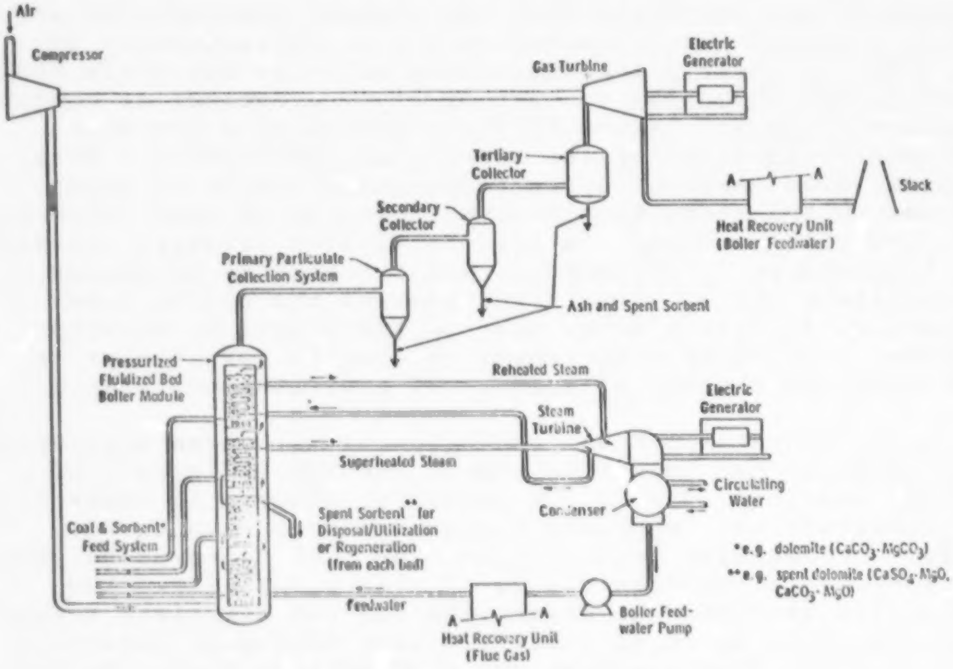
Air
Compressor
Gas Turbine
Electric Generator
Tertiary Collector
Secondary Collector
Primary Particulate Collection System
A
A
Stack
Heat Recovery Unit (Boiler Feedwater)
Ash and Spent Sorbent
Pressurized Fluidized Bed Boiler Module
Reheated Steam
Electric Generator
Steam Turbine
Superheated Steam
Circulating Water
Condenser
Coal & Sorbent* Feed System
Spent Sorbent** for Disposal/Utilization or Regeneration (from each bed)
Feedwater
A
A
Heat Recovery Unit (Flue Gas)
Boiler Feed-water Pump
*e.g. dolomite ($CaCO_3 \cdot MgCO_3$)
**e.g. spent dolomite ($CaSO_4 \cdot MgO$, $CaCO_3 \cdot MgO$)

Figure 1 - Pressurized fluidized bed boiler - power plant schematic

both the steam tubes immersed in the fluidized bed and by the heat convertor boiler of the gas turbine. The heat of combustion is removed from the bed to run directly to the gas-turbine. Typically such a fluidized bed uses crushed coal of about 1/4" diameter. Most of the fly-ash particulate particles are rather large, say 100μm; however, some smaller particles are also produced which must be removed from the flue gas to avoid the attack of gas-turbine blades by such corrosive particulates as sodium salts.

Cyclones are used to effectively remove large particles. For particles in the range of 2-20μm diameter non-conventional devices are suitable for this application. Filters are difficult to use because of the high temperature, corrosive environment and because of the inherent high-pressure drop across the filters. The electrostatic precipitator is not applicable because under the high pressure (10-20 Atm in the fluidized-bed boiler) it is difficult to produce the corona discharge without generating excessive electric arcs. Again, the use of an acoustic agglomerator, operating either in standing-wave or traveling-wave mode, appears very attractive.

SOLID-SEED EXTRACTIONS FROM MHD GENERATORS

Because of the increased cost and reduced availability of fossil fuels, a strong incentive now exists to substantially increase the electric power plant efficiency which is currently running between 33-40% depending on whether it is a nuclear or conventional fossil-fuel plant. Since 1960 new plants have not done appreciably better in terms of the power plant efficiency. This comes primarily from inherent upper temperature limits of about 1050°F imposed on the steam-turbine cycle because of some inherent thermal and material problems. Alloy steels lose strength rapidly above this temperature. Furthermore, water begins to decompose at this temperature and it attacks and weakens the boiler tube walls. Consequently unless a new thermodynamic unit is developed we have reached a point of diminishing returns in our attempt to further increase the thermal efficiency of central power plants.

Magnetohydrodynamic (MHD) topping units have been developed as a way to increase the peak temperature and the efficiency of electric power stations [23-25]. A schematic diagram is shown in Fig. 2 to illustrate the important components that have to be developed for a commercial unit of this type [23]. One of the key technical problems in the development of such a coal-fired MHD powered generator involved the separation and the efficient removal of the seed material, which is usually either sodium or potassium, before releasing the flue gas into the atmosphere. Again, because of the high temperature and the excessively corrosive environment,

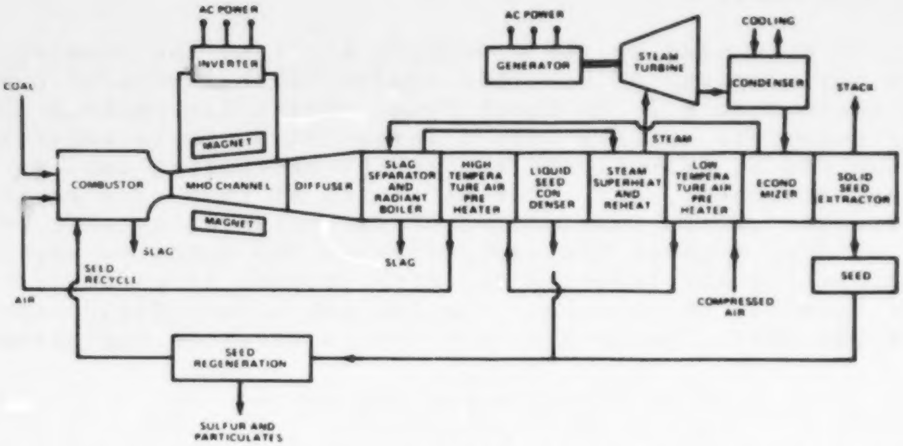
AC POWER
INVERTER
COAL
MAGNET
COMBUSTOR
MHD CHANNEL
DIFFUSER
MAGNET
SLAG
SEED RECYCLE
AIR
AC POWER
GENERATOR
STEAM TURBINE
COOLING
CONDENSER
STACK
STEAM
SLAG SEPARATOR AND RADIANT BOILER
HIGH TEMPERA TURE AIR PRE HEATER
LIQUID SEED CON DENSER
STEAM SUPERHEAT AND REHEAT
LOW TEMPERA TURE AIR PRE HEATER
ECONO MIZER
SOLID SEED EXTRACTOR
SLAG
SEED
COMPRESSED AIR
SEED REGENERATION
SULFUR AND PARTICULATES

Fig.2 Schematic diagram of important MHD generator components (Taken from ref. [23])

conventional devices are difficult to use in this situation. A recent study shows that cyclones can be used to remove most of the large particles (diameter larger than 10µm). New techniques must be found to collect particles in smaller sizes. Particles are mostly spherical because of the very high operating temperature at which MHD generators perform, typically between 2000-2500°C. Figure 3 shows typical agglomerates collected from cyclones from a MHD generator [23].

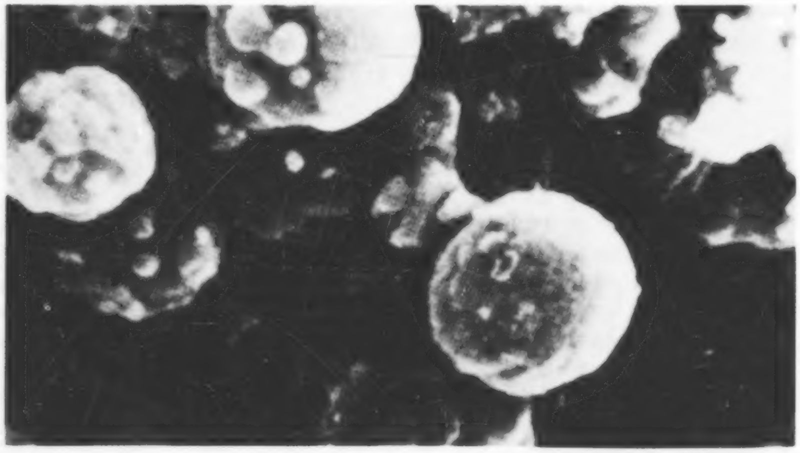

Fig. 3 Typical agglomerates collected from cyclones from a MHD generator (Reference [23])

THEORETICAL ASPECTS

As discussed in the preceding section, the interest in various applications of acoustic agglomerators concerns particles in the range from 0.2- to about 20μm. Particles smaller than 0.2μm usually coagulate quickly into a larger size; while particles larger than 20μm can be effectively removed by mechanical devices such as cyclones, scrubbers, etc. For particles of this size region acoustic agglomeration offers some unique characteristics as will be explained later. Figure 4 shows the variations of the relative amplitudes of the oscillating particles to gas with respect to particle radius for various acoustic frequencies. On the top of the figure the oscillation of the particles is also plotted in terms of the parameter τ/T.

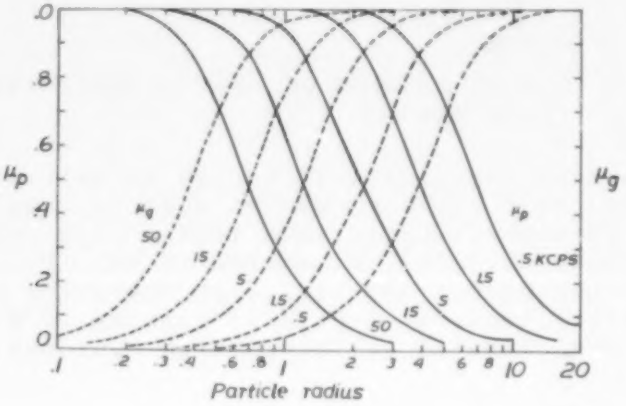

Fig. 4 Variation of relative particle amplitudes vs. particle radius for various particle frequencies

It is seen that the particles vibrate virtually with the same amplitude as the carrier gas at low frequencies. As the sonic frequencies increase the inertia of the particle becomes increasingly evident until eventually a frequency is reached in which the particles remain virtually stationary despite the intense vibration of the carrier gas. In the intermediate frequencies the particles precipitate in the vibrations to varying degrees, depending upon the sizes. For example, if the frequency of the acoustic field is 20 Kc/sec, particles with diameters smaller than 2μm vibrate with nearly full amplitude ($X_p/X_g = 1$), while particles with diameter larger than 4μm are virtually stable ($X_p/X_g = 0$). This differential motion of vibrating particles leads to the agglomeration of particles.

Since the inertia of a particle can be conveniently expressed in terms of τ, which can be expressed as:

$$\tau = \frac{2}{9}\ \frac{\rho_p}{\eta}\ a^2 \qquad (1)$$

where ρ_p is the particle density, η is the dynamic viscosity of the carrier gas and a is the particle radius. $T = 2\pi/\omega = 1/f$ where ω and f are the angular velocity and frequency of the acoustic field, respectively. τ/T characterizes the relative effectiveness of the inertial force during one oscillating cycle. When τ/T >> 1 the inertia is too large for the particle to respond to the acoustis oscillation. Thus, the particle can be considered to be relatively stationary. When τ/T << 1.0, the inertia is negligible and the particle is oscillating together with the carrier gas. The degree to which a particle is carried along with the oscillating motion is specified by an entrainment factor, μ_p which is defined by:

$$\mu_p = \frac{X_p}{X_g} = [1 + 4\pi^2(\tau/T)^2]^2 = \frac{U_p}{U_g} \qquad (2)$$

where X_p and X_g are the oscillating amplitudes of particles and the carrier gas, respectively. Equation (2) is plotted in Fig.4.

Owing to the differential motions, smaller particles are agglomerated into larger ones and the time rate of small particle concentration can be written as:

$$\frac{dn_2}{dt} = -K_a n_2 \qquad (3)$$

Where n_2 is the small particle concentration. Integrating Eq.(3) and using the initial conditions n_{20} at t = 0, we have

$$n_2 = n_{20}\, e^{-K_a t} \qquad (4)$$

where K_a is defined as the acoustic agglomeration constant given by

$$K_a = \alpha \, \varepsilon \, \pi(a_1 + a_2)^2 (2X_p) n_1 f \qquad (5)$$

where α is a filling-up factor, n_1 is the concentration of large particles which act as collectors of small particles, X_p is defined in Eq.(2), a_1 and a_2 are the radius of large and small particles, respectively. ε is defined as the particles collision efficiency. To understand the acoustic agglomeration mechanisms it is necessary to describe briefly the theoretical predictions of the values of ε. The particles of diameter range 0.2-20μm, the calculation of ε may be divided into two subregions. The first region is for particles of diameter range from 0.2 to about 2μm. For these particles the predominant mechanism for acoustic agglomeration is the so-called ortho-kinetic interaction. Basically, orthokinetic interaction occurs between two particles of different sizes which are moving at different velocities according to Eq.(2). Such differential motion causes the two particles to collide with each other, leading to agglomeration into one larger particle. For this interaction the particle efficiency ε can be expressed as [26,27]:

$$\varepsilon = K_e^2 (K_e + 0.25)^{-2} \qquad (6)$$

where $$K_e = \frac{2\mu_g a_1}{g} (\Omega/\rho_g)^{1/2}$$

Typical results of Eq.(6) are shown in Fig. 5. For particles of diameter range from 2-20 the predominant mechanism for acoustic agglomeration is the so-called hydrodynamic interaction. In this case particles interact with each other due to the hydrodynamic interaction forces resulting from the disturbance of the flow field of one particle to another. It is important to point out that the orthokinetic interaction is not important in this region because particles larger than 20μm seldom exist in any of the applications as discussed in this paper. For this mechanism the particle collision efficiency ε can be determined by using the following expression for y_c

$$y_c = 1 + I - \frac{B}{(I^6 + p_1^6)^{1/6}} - \frac{B}{((1-I)^{1.5} + p_2^{1.5})^{2/3}}$$

$$\text{where } B = \frac{1.587}{a_e} + \frac{32.73}{a_e^2} + \frac{344}{a_e^2(20/a_e)^{1.56}} \exp - \frac{a_e-10}{15} \sin \frac{a_e-10}{63} \qquad (7)$$

$$a_e = (2\mu_g/\tau_1 g)^{1/3}(\Omega/\rho_g)^{1/6} a_e$$

p_1 and p_2 are constants such that $y_c = 0$ when $I = 0$ and $y_c^2 = 1/4$ when $I = 1$ and are determined by iteration for each a_e.

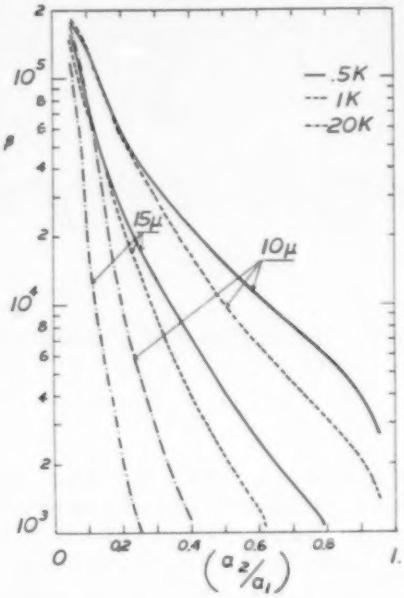

Fig. 5 β (ratio of acoustic agglomeration rate to the agglomeration rate without acoustic field) vs. particle-radius ratio for various acoustic frequencies (in kHz) for large particle radius (μm)

Typical results of Eq.(7) are shown in Fig. 6. For both types of interactions a very large increase in the agglomeration rate as compared with the conventional gravitational coagulation can be expected. However, it is important to point out that the theoretical prediction in both Figs. 5 and 6 are likely to be too optimistic since in both figures the filling-up factor α (as shown in Eq.5) is assumed to be one. In practical cases α is less than one. The actual value depends mostly on the acoustic turbulent strength. No information, neither experimental nor theoretical, is available on α at the present time. It should be noted also that for control of small particles (diameters less than 2μm) a high frequency acoustic field is much more effective; while for large particles (diameter between 2-20 m), a low frequency field is much more effective.

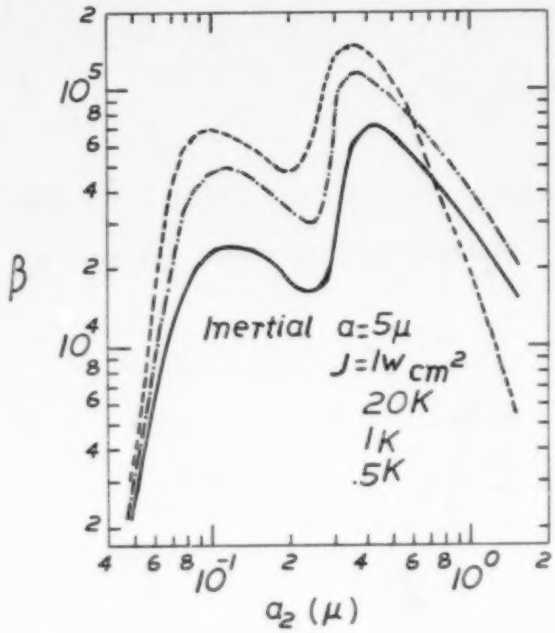

Fig. 6 β vs. small particle radius for various acoustic frequencies ($a = 5\mu$, $j = 1 w/cm^2$)

The rate of reduction, as indicated in Eq.(3) depends on the square of particle concentration. Thus the higher the particle concentration the more effective is the acoustic agglomeration. For the LMFBR applications, typical mass loading is high (range from 1-30g/m^3), thus acoustic agglomeration is expected to be very effective. For other applications in which the particle number concentrations are not as high as in the case of MHD and fluidized bed boilers, it may be necessary to add a mist to bring the particle loading to a satisfactory level. However, the mist must be chemically and thermally stable under the high temperature and high pressure environment, such that the mist droplets will be removed completely by the cyclone after leaving the acoustic agglomerate. The required acoustic intensity depends on the requirements on particle agglomeration time, since the acoustic coagulation constant K_a is directly proportional to the square root of acoustic intensity.

ACKNOWLEDGEMENT

This work was supported in part by EPRI and NSF.

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WRITTEN DISCUSSION

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Argonne National Laboratory's interest in controlling particulate emissions with acoustic agglomerators arises from its participation in a research and development program designed to develop the concept of fluidized bed combustion (FBC) and to eventually demonstrate on a plant scale that the process is economical and meets pollution standards for stationary power sources. Simply stated, fluidized bed combustion is a process in which coal is burned with excess air in a fluidized bed of limestone or dolomite at temperatures of 800 to 950°C. The heat of combustion is used to generate steam for process use or to generate electrical power. Fluidized bed combustion, which results in both reduced SO_2 emissions (by reaction of the SO_2 with CaO in the fluidized bed) and reduced NO_x emissions (by reason of the low combustion temperatures), is potentially both energy efficient and economically attractive.

In advanced concepts of FBC, the combustor is operated at elevated pressure (e.g., 10 atm). Energy is extracted from the process by raising steam in boiler tubes and by expanding the flue gas through a gas turbine to take advantage of the potentially higher thermal efficiencies offered by a gas turbine-steam turbine combined cycle. Use of a gas turbine, however, requires that the particulate loading of the high temperature-high pressure flue gas from the pressurized FBC (PFBC) be reduced to very low levels. Estimates of acceptable dust loading for turbine operation range from 0.05 to 0.0002 grains/scf.

Until recently, the problem of high temperature-high pressure flue gas cleaning (particularly as it relates to PFBC) was not adequately addressed. At present, however, a fairly intensive effort is being made to develop viable techniques for sufficient cleaning of flue gas from a pressurized fluidized bed combustor to allow its use in reliable gas turbine operation.

Currently, ANL is conducting basic supportive studies in fluidized bed combustion. One task in our program is to test and evaluate novel concepts for high temperature-high pressure flue gas cleaning in our process development facility, which includes a 6-in.-dia. fluidized bed combustor which can be operated at pressures up to 10 atm. Initially, a review was made of the existing and developing particle-gas separation technologies to identify concepts which would be amenable to the high temperature-high pressure requirements of pressurized FBC.[1] One area identified for investigation was acoustic agglomeration. As a result, we entered into a working relationship with Dr. David Scott of the University of Toronto, who is also presenting a paper on acoustic agglomeration at this workshop. It is our intention to investigate the effect of acoustic dust conditioning on the efficiency of particulate removal devices in the flue gas system of the ANL experimental combustor.

Although the agglomerating influence of an acoustic field on an aerosol has been known for slightly over 100 years, the development of acoustic agglomeration (AA) as an aid in the control of particulate aerosols has passed through cycles varying from enthusiasm to almost total disinterest. When the electrostatic precipitator won the performance/cost war in about 1952, interest in acoustic agglomerators in the U.S. essentially went to zero. At the same time, however, interest in AA was generated in the U.S.S.R., Hungary, Poland, and Japan. This interest in foreign countries has continued to the present. As a result, the U.S. has been left behind in both developing this technology and maintaining the necessary scientific interest.

Some of the factors which led to the premature disinterest in AA were: (1) a poor understanding of the aerosol mechanics of AA, (2) less than optimum configurations in the designs originally developed for commercial application, (3) a demonstrated superiority of electrostatic precipitators, (4) concern over the high power requirements for the generation of the acoustic field, and (5) potentially high capital cost for acoustic agglomeration systems.

In the paper by Shaw and Wegrzyn, the argument is made that sonic agglomerators cannot compete in applications where conventional devices already have been successfully used, but that they have some characteristics which make them uniquely suitable for certain special applications. The phrase "uniquely suitable" must not be interpreted to imply a priori that sonic agglomeration is necessarily the most suitable control technique for these special applications. In the example of the LMFBR sodium-fire aerosol, although the acoustic control technique was ranked first in effectiveness, it was ranked ninth in overall preference due to other considerations. Also although acoustic agglomeration appears attractive in controlling emissions from a fluidized-bed boiler to protect a gas turbine in a gas turbine-steam turbine combined cycle, a number of alternative (and generally more highly regarded) control technologies are being investigated. Granular-bed filters and ceramic filters (either fabric or porous sintered) are the more commonly mentioned alternatives being considered for this application.

It is interesting to note in the ranking of control methods by criteria for the control of LMFBR sodium-fire aerosols (Table 2 in paper by Shaw and Wegrzyn), one of the problems facing acoustic agglomeration was that of credibility. This same problem of credibility exists in the application of acoustic dust conditioning to PFBC applications. The problem of credibility arises mainly from the lack of experimental correlations or experimentally verified theory which can be used to predict either performance or cost in a given application.

The credibility problem is really not too surprising in view of the complexity of the equations which govern the aerosol mechanics and the fact that the experimental data on acoustic agglomeration are for the most part of no use for evaluating theory. If AA is to gain credibility, both the performance and the energy requirements have to be established (either by experiment or from reliable theory) for the particular application being considered and a performance-cost comparison made against competing technologies.

A good argument for determining energy requirements can be made in the case of particulate control from pressurized fluidized-bed combustors. For atmospheric fluidized bed combustion, an overall plant efficiency of 36-39 percent is estimated. In atmospheric operation, the flue gas can be adequately cleaned using conventional technology. First generation pressurized FBC systems will have an estimated plant efficiency of 40-43 percent. The small but significant gain in plant efficiency for pressurized systems is at the cost of considerably more complex development problems (of which cleaning of high temperature-high pressure flue gas is one). If cleaning of the high temperature-high pressure flue gas to acceptable particle levels for turbine operation by AA or any other method is excessively energy-intensive, the PFBC loses much of its advantage over the atmospheric concept.

With reference to the theoretical aspects of acoustic agglomeration as presented by Shaw and Wegrzyn, the following comments are offered. First, the authors are to be commended for their efforts to identify the controlling mechanisms in AA and to experimentally verify the respective governing equations for these mechanisms. It should be emphasized, however, that the agglomeration mechanisms considered by Shaw and Wegrzyn (i.e., orthokinetic vibration and hydrodynamic attraction) are but two of approximately nine mechanisms known to have some effect on the agglomeration kinetics. The problem is sufficiently complex that the mechanisms of AA may depend upon the nature of the aerosol, the acoustic field, and the system geometry. Thus, the conclusions of a particular study may be seemingly contradicted by the results of other experiments when both sets of data are equally valid but depict different mechanisms under different circumstances. The statements, therefore, by Shaw and Wegrzyn that the predominant mechanism is acoustic agglomeration for particles of diameter range 0.2 to 2 μm is ortho-kinetic interaction and the predominant mechanism for particles of diameter range 2 to 20 μm is hydrodynamic interaction should not be accepted without justification as general observations which can be applied to all acoustic-aerosol interactions.

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OPEN DISCUSSION

Shaw: I agree with what Bill said about the nine mechanisms he mentioned. The other seven are mostly associated with the turbulence. The parakinetic is also what I term as the filling up of the space. So I agree with what you are saying on the nine mechanisms. However, I don't think you can really identify the way we are approaching the problem on the mechanism beyond these two. We have already identified the linear case; we are also looking at the affect of turbulence overall. You have to define some parameter in terms of acoustic turbulence which I think will cover other mechanisms such as the parakinetic and so on, because there is no way to separate the parakinetic from the turbulence under high intensity. They are interrelated.

We have done separate things in a nonlinear case, theoretically and a few experiments under high intensity; but we are far away from matching the theory and experiment.

Yang: Bill, is the work that you are just finishing available?

Swift: No, it is in a state of being published, both as an Argonne report and an NTIS document.

Oder: Could you comment on the strength of coagules that are formed by the agglomeration process? I gather that they are brought together by turbulent acoustic forces and they settle out in rather undisturbed environments. Suppose, instead, that you try to extract these particulates under turbulent conditions, what is it that holds them together?

Shaw: That is a very good question. So far we have only done experiments with particles of a very small size, primarily for the application of fine particulate control which is supported by EPRI. We have not done any work on the high concentration yet. We are expecting a contract from ERDA on the high intensity agglomeration of large particles for sodium aerosol. The small particle sticking is not a problem. That comes when large particles, sticking together, break apart. In that case, some kind of salient condition must exist so that they stick together. This is something that we are just still speculating, we have no experimental data to show. The breaking up by the acoustics will be important only when you have very large particles.

Oder: I am sorry, I still didn't hear what it is that makes them stick together.

Shaw: Surface molecular forces. Just like when you have a particle on an impactor, it is the same force holding a particle to an impactors plate as when the particles are sticking together.

Helfritch: You mentioned that, most often, the use of agglomeration techniques applies to cyclones. There are two questions that have been brought up about the success of that application. The apparent specific gravity of the agglomerates would be quite low. Secondly, they may break up. I was wondering if it had actually been tried in a cyclone.

Shaw: No, we have not tried it. The only reason mentioning cyclones is that it can only be used in applications like the Argonne system where they have a cyclone and want to work it with something else. Whenever you have a situation where the pressure drops and pumping is a problem, the cyclone seems to be the answer. It requires very little pumping power.

Helfritch: You mentioned that a high pressure drop across the filter is part of the reason for projecting that technique. However, I would submit that the pressure drop across the filter is typically on the order of pressure drop across the cyclones. I cannot accept that argument.

Shaw: If that is the case, then I am wrong. When I talked to Argonne people, they told me that they cannot use filters. My impression is that the problem with filters are high temperatures, the other is a pressure drop problem.

Helfritch: Finally related to that is the fact that high temperature, and pressures adversely affect performances of cyclones. That is because of the viscosity relationship.

Swift: By acoustically treating the aerosol you can increase the efficiency of the cyclone over that of the untreated aerosol even though you have high temperatures and pressures.

Cooper: The problem with the sodium aerosol is that it clogs the filter in a way that is very unusual and very difficult. So you are not talking about the usual few inches of pressure drop and any reasonable phase velocity; you are talking about much higher than that. Second, there is a difference in degrading the performance of the cyclone by changing its collection efficiency cut point by the higher temperature and pressure. And indeed, the degradation of the filter is where it just falls apart.

Grassel: It seems to me that the work that is going on now, is just repeating work that was done a number of years ago by the Bureau of Mines and other places during the war. Yet it does not address itself to the practical problem of implementing sonics as a method of treating aerosols. It does not address itself to the problem of true output, sound level and size. You are going to have to give up one of those for the others; sound level is what you are attempting to compromise. You are attempting to use sound level as a method of accomplishing size and true output. This represents itself in power. As the power goes in, you have two practical problems of application: containment of the sound level and dissipating it.

The power necessary to agglomerate particles is very low. The power necessary to generate a field of sufficient intensity to collect those particles or to agglomerate those particles is very high.

Shaw: What you are saying is the power attenuation in the particle is small.

Grassel: But the power put into the system is very large.

Shaw: You must obtain a level of 160 dB. That is precisely the reason why we think a standing wave kind of thing would be good for fine particle control.

In response to your comments, we are doing something like the whistle experiments in the 1950's. We have to go through this phase to reactivate the field. The thing we are doing different nowadays is that the particle sizing techniques are much more improved than the wartime. I do not agree that others have done the same thing. No one before has measured size distribution change in an acoustic field. The only other related work has been done by a Russian who sampled the size by putting in something to collect the particles by sedimentation.

What we are doing in our laboratory is trying to bring us to large scale testing to give us credibility and demonstrate to people that it will work. If you do it right it will work. The problem is how well it will work. This is the thing we are driving at. We have to go through the large scale testing which is nothing new. I emphasize that the state of our siren is 1949. But we are applying the new sizing technology to study this problem.

Benarie: Let me take just a minute for this problem of the cyclone. Perhaps it would simplify concepts if I explained that we have vortex agglomeration for both cases: (1) a cyclone with a hardware around the vortex and (2) when there is no hardware around it. It is a turbulent agglomeration.

Shaw: In no way do we think the acoustic cyclone works the best. In fact, from the data, we feel that some kind of acoustic thing working together with a scrubber, with a very fine mist, for certain applications is best. With high temperatures you have trouble maintaining that mist.

Scott: All the work that was done at the Ontario Research Foundation was with a cyclone. So the improvements noted, from 35 and 40% up to 75 and 80%, were done in cyclones. It is clear that we want to go that way again. Second, on the power, that was what the whole project was about. It was to determine the cost of the power. Our evaluation has shown that if you use your fuel evaluation for the pulse jet and take the cost of the fuel, then the power requirements are way down for scrubbers. We are talking about 25¢/ACFM/year for power. That is for a pulse jet, which in that case was designed with an aerodynamic valve. I think I simply must not let the thing go without saying that cyclones have been used. Secondly, we have concentrated on power. That was really the whole topic of our work. Lastly, as far as the Bureau of Mines is concerned, I do not think we are repeating very much. I would be happy to see those areas where you think we are. I think we are doing things that are quite different.

Cooper: Generally, surprisingly little power is required to remove particles from the gas stream if you just take the force on to particles times the distance they are to be moved. In some sense we ought to keep that in mind.

That is one of the advantages of electrostatic and magnetic separation as opposed to other methods which operate on the gas which often has very much more mass than the particles. There is an inherent advantage to those systems which apply the force directly to the particle rather than through the medium of the gas to the particle.

EFFECTS OF ELECTRIC AND ACOUSTIC FIELDS ON PARTICLE COLLISION RATES IN AEROCOLLOIDAL SUSPENSIONS

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ABSTRACT

An experimental investigation is described that involves the use of an a.c. electric field and a traveling sound wave field to increase the collision rate between submicron particles suspended in air. A brief discussion of the theoretical collision rates in the free molecule regime is also included for the cases where each field is applied separately. Velocity profiles and values of the turbulent intensity in the continuous flow coagulation tube are presented and discussed. In addition, the research plan is briefly outlined.

I. INTRODUCTION

The work reported here is based on an experimental project that is underway to investigate the effects of an a.c. electric field and a traveling wave sound field on the collision rates between suspended particles with radii less than about 0.3 μm. We have chosen to work with particles in the submicron range because of the compelling need to find ways to effectively remove these particles from man-made suspensions. Although we are not dealing with the direct removal of the particles from the parent gas, the work discussed here involves the use of two external fields to "condition" the suspension by inducing particle growth of the fine particulate matter through agglomeration or coalescence. Following the "conditioning" process, the suspension is envisioned to be passed through a downstream and conventional removal device where the resulting agglomerates are to be removed. The induced growth is achieved by using the external fields to selectively increase the relative motion and therefore, the probability of collisions between the suspended particles.

The motion of the particles under the influence of an a.c. electric field and a traveling sound wave may be demonstrated in the following way. Consider a charged particle with a charge q and mobility B suspended in a gas. The velocity of the particle in the presence of an a.c. electric field is

$$\underset{\sim}{v} = q\ B\ \underset{\sim}{E} \sin \omega_e t,$$

where ω_e is the angular frequency of the field and $\underset{\sim}{E}$ is the electric field vector. Here we assume that the product of the angular frequency and the particle relaxation time, τ, is much less than unity, i.e., $\omega_e \tau \ll 1$. Now consider that a continuous train of traveling planar sound waves passes through the suspension. The velocity of the particle in this case is [1]

$$\underset{\sim}{v} = \frac{U_g}{(1 + \omega_a^2 \tau^2)^{1/2}} \quad [\sin (\omega_a t - \psi)]\underset{\sim}{\hat{n}} \quad ,$$

where ω_a is the angular frequency of the wave, U_g is the magnitude of the oscillating velocity of the gas molecules, ψ is the phase lag between the particle and the oscillating gas molecules, and $\underset{\sim}{\hat{n}}$ is a unit vector in the direction of propagation. The magnitude of the particle velocity,

$$U_g/(1 + \omega_a^2 \tau^2)^{1/2} \quad ,$$

and the phase lag, ψ, are dependent on the product $\omega_a \tau$, where τ is dependent on the size and mass of the particle. For $\omega_a \tau \ll 1$, the particle oscillates at nearly the same speed as the gas molecules and with very little phase difference. On the other hand, for $\omega_a \tau \gg 1$, the particle is virtually stationary. (For two different particles, one with $\omega_a \tau \ll 1$ and one with $\omega_a \tau \gg 1$, there is a large difference in the oscillating speeds. Such a differential motion is the basis for orthokinetic coagulation.)

By superimposing the two fields, the resultant motion of the particle is a function of the two field frequencies and magnitudes, the directions of the two fields, and of the particle parameters, which include charge, size, and mass. It is this resultant motion of many particles submerged in these two fields which can lead to an increase in the particle-particle collision rate that is of interest here.

It should be noted that, whereas acoustic coagulation generally requires high frequency sources (i.e., on the order of 10-30 kHz) to set up the differential motion in a suspension of fine particles, the present two-field scheme utilizes full entrainment of all the particles and hence lower sound frequencies. This feature has some sound transducer and power supply advantages. However, this scheme also requires that the particles be charged, which necessitates the addition of some type of charging stage in the system. It is also noted that a traveling wave sound field, as opposed to a standing wave field, is considered here following the work of Scott [2].

For this case of combining an electric field and a sound field to promote an increase in particle-particle collisions, there is no known theoretical expression that relates the collision rate to the field and aerosol parameters. However, past work involving derivations of the collision rates for electric field enhanced coagulation and for acoustically induced coagulation can be used as a basis for considering a new model for the case where both fields are applied.

A. Theoretical Background

Using the particle Knudsen number to identify the usual aerosol size regimes (see Hidy and Brock [3], for example), suspensions of particles at standard conditions with radii in the approximate range $0.01 < R < 1$ μm are classified as transition regime aerosols. (Continuum regime aerosols have particle radii larger than about 1 μm and free molecule regime aerosols have particle radii less than about 0.01 μm.) The particle radii of interest here are in the transition regime, and herein lies a problem for treating the collision dynamics of these particles. Whereas, continuum mechanics is used in treating the dynamics of continuum aerosols and kinetic theory is used in the free molecule regime, the transition regime is without a single, unified theoretical base. Extensions of continuum regime theories and of free molecule theories into the transition regime for the thermal collision rate are available [3,4]. Experimental results show that in some cases these extensions are reasonably accurate [4]. However, extensions do not exist for collision rates involving electric or acoustic fields. Expressions are available for the collision rate as a function of the field quantities for a sound field in the continuum regime (see Mednikov [1]) and for an electric field in the free molecule regime (see Brock and Hidy [5]). Rather than try patching together existing theories and then extending the results into the transition regime, we have taken the approach of using a free molecule theory to construct an expression for the collision rate due to these external fields. This approach, which is discussed in the next major section, emphasizes the diffusional processes and also provides a framework for dealing with the basic collisional dynamics of the suspended particles in the presence of the two fields. There is evidence that diffusional effects are significant in sonic agglomeration of particles with radii below about 1 μm [6]. This fact is reason enough to justify the development and examination of this free molecule approach.

B. Summary of Past Experimental Work

There is very little experimental data available on the collision rates of suspended submicron sized particles subjected to electric or acoustic fields. In the transition regime, Mednikov [1] reports some experimental results of acoustic coagulation in DOP fogs with mean particle diameters of about 0.28 μm. The coagulation constant was found to be directly proportional to the product of the sound pressure and the time that the sound field was applied to the fog. Details of the experiment are vague and the accuracy of the results is unknown. The work in the past has involved aerosols that were not very monodisperse, and hence the reported collision rates are not very size sensitive. In addition, many of the earlier studies involved time decay measurements in a closed chamber and were subject to significant diffusion loss effects to the chamber walls as the coagulation process evolved and the samples were extracted for size analysis [6]. Currently, there are several experimental investigations underway that should contribute significantly to the understanding of aerosol-acoustic field coupling. Much of this current work is possible because of significant improvements in the equipment available to generate and analyze aerosols. The experimental investigation reported here is a product of this improved technology.

C. Paper Organization

The next section, which is Section II, is a summary of a first order derivation using kinetic theory to determine the particle-particle collision rate as a function of the aerosol parameters and the field frequency and magnitude for an a.c. electric field. The derivation of the collision rate for a traveling sound wave is also summarized. The expressions for these rates, along with the known expressions discussed earlier, are to be tested during the experimental investigation, which is underway and which is described in the third section of the paper. Section III also contains the results of velocity and turbulence measurements taken in the coagulation tube. In addition, the experimental plan is briefly discussed. Section IV consists of a summary of the paper.

II. COLLISION RATES IN THE FREE MOLECULE REGIME

As mentioned before, an expression for the collision rate between particles in an aerocolloidal suspension subjected to a combined electric field and traveling sound field does not exist. It is our desire to construct a simplified model to describe the particle-particle collisions in this case. Because diffusional effects are important in the collisional processes involving submicron particles [6], it was decided to approach the problem using a free molecule regime model. However, before solving the more complicated combined field problem, it was decided to derive the expressions for the collision rates for each field separately to better understand the effect that each field has on particle collisions. These derivations have been completed recently and the following is a summary of that work. The derived rates are to be checked with experimentally determined rates during the experimental investigation which is underway and which is discussed in Section III of this paper.

A. General Approach to Deriving the Collision Rate in the Free Molecule Regime

In the free molecule regime, we treat the aerocolloidal system as a dilute mixture of gas molecules and two separate, but homogeneously mixed monodisperse aerosol species with particle radii R_i and R_j and concentrations n_i and n_j, respectively. For each species, the particles are considered to be spherical and the number density of either particle species is much less than the number density of the gas molecules, while the mass per particle of either species is greater than the mass of a gas molecule. (This is equivalent to a very low mass loading.) In this case the three governing Boltzmann equations (i.e., one for each type of aerosol species and one for the gas molecules) are uncoupled and the equations for each aerosol species are identical, except for a difference in subscripts. An approximation technique is then used to solve one of the aerosol equations for the particle distribution function, say f_i, after assuming the form of the distribution function for the suspending gas. (The second species distribution function, f_j, is the same, except for a change of subscripts.) An expression for the collision rate is then determined by evaluating the collision integral of the products of the two particle distribution functions for a specified particle-particle force law. This approach has been discussed elsewhere (see Scholz and Barber [8], for example) and will not be presented here.

B. Collision Rate for an Aerosol with an a.c. Electric Field

In this case we assume that the dilute mixture is spatially uniform and that the particles in each species have an average charge per particle of q_i and q_j, respectively. The acceleration due to the externally applied electric field is

$$\underline{a}_r = \frac{q_r}{m_r} (E \cos \omega_e t) \underline{\hat{z}} \quad , \qquad (r = i,j)$$

where m_r is the particle mass of each aerosol species and $E \cos \omega_e t$ is the magnitude of the z directed, time varying electric field. In addition, the gas is assumed to obey a Maxwellian velocity distribution function.

Following the procedure outlined by Scholz and Barber [8], the Boltzmann equation for the i^{th} aerosol species may be solved by an approximation scheme using an expansion with a weighting function equal to some zero-order approximation of f_i. In this work the zero-order function is taken as the Maxwellian distribution function. In addition, the particle-gas collisions are assumed to obey an inverse fifth power central force law (i.e., the Maxwell force law). Using this approach, the approximate particle distribution function for a three term expansion is

$$f_i \doteq D_o \left\{1 + \beta \,[\sin(\omega_e t - \phi)]\, v_{iz} + \frac{1}{2} \beta^2 [\sin^2(\omega_e t - \phi)](v_{iz}^2 - \frac{kT}{m_i})\right\} (f_i)_M ,$$

where
$$\beta = \frac{q_i E}{kT} (\omega_e \sin \phi - C_1 \cos\phi)^{-1}, \qquad \phi = - \tan^{-1}(\omega_e / C_1),$$

$$C_1 = 2\pi K^{1/2} \frac{1}{m_i} \left(\frac{m_i m_g}{m_i + m_g}\right)^{1/2} n_g A_1 \quad ,$$

and where $(f_i)_M$ is the Maxwellian function for the i^{th} aerosol species. In these expressions n_i and m_i are the aerosol particle concentration and mass for species i, n_g and m_g are similar quantities for the gas molecules, v_{iz} is the particle speed in the z direction (and is a variable), k is the Boltzmann constant, T is the absolute temperature of the suspension, A_1 is a collisional constant for the particle-gas collisions and equals 0.422, K is the inverse fifth power force constant and D_o is an arbitrary constant. As mentioned earlier, the expression for f_j is identical except the i subscripts are replaced by j.

It should be noted that this expression for f_i differs from the distribution function used by Brock and Hidy [5] to determine an expression for the electric field induced collision rate. In that work, they expressed the effect of the electric field as an equivalent drift term. This term was then included in the argument of the exponential velocity term of a Maxwellian velocity distribution function used to construct f_i. The approach here is more fundamental for we have accounted directly for the effects of gas-particle collisions in the collision integral when we solved the Boltzmann equation for f_i.

As mentioned previously, the collision rate is determined by evaluating the collision integral of the products of the i and j particle distribution functions for a specified particle-particle force law. Following Hidy and Brock [3], the specified force law for the particle-particle collisions is based on a hard sphere collision model with a Debye-Hückel shielded electrostatic potential acting between the colliding and charged spheres. The collision rate, L_E, for the special case where the two species are the same size (i.e., $R_i = R_j$) and have the same, but opposite average charge per particle is

$$L_E = D_o^2\left[1 + \frac{kT}{6m_i}\,\beta^2\sin^2(\omega_e t - \phi) - \frac{1}{4}\left(\frac{kT}{6m_i}\right)^2 \beta^4\sin^4(\omega_e t - \phi)\right] L_T,$$

where L_T is the self-thermal collision rate and is

$$L_T = 8\sqrt{\pi}\;\; \eta R_i^2 n_i^2\, \Omega \left(\frac{kT}{m_i}\right)^{1/2} \quad .$$

In this expression η is the fraction of particles that stick upon contact and Ω is the charged particle interaction parameter and is

$$\Omega = 1 + \frac{q_i^2 \exp(-\,2R_i/\lambda_D)}{16\pi\ \varepsilon\ kTR_i} \quad ,$$

where ε is the dielectric constant of the aerosol, and

$$\lambda_D = \left(\frac{\varepsilon kT}{8\pi n_i q_i^2}\right)^{1/2}$$

is the Debye radius. This case where $R_i = R_j$ and $q_i = -q_j$ corresponds to a quasi-neutral, monodisperse aerosol and might represent an initially neutral, monodisperse aerosol that is passed through a bipolar charger in which complete charging takes place.

C. Collision Rate for an Aerosol with a Traveling Sound Wave

In this case we consider planar sound waves propagating in the positive z direction with a speed of c and an angular frequency of ω_a. The gas is assumed to obey the Wang-Chang and Uhlenbeck [9] distribution function for small amplitude sound waves propagating in a monatomic gas. That is,

$$f_g \simeq \left[1 + \alpha_{oo}\left(\frac{v_{gz}}{c} + \frac{5v_g^2}{3c^2}\right)\sin\omega_a\left(t - \frac{z}{c}\right)\right](f_g)_M \; .$$

In this expression α_{oo} is an arbitrary constant, the g subscript refers to the gas, and $(f_g)_M$ is a Maxwellian distribution for the gas. The arbitrary constant, α_{oo}, is evaluated here by setting the z component of the average speed of the gas at the piston face (i.e., at $z = 0$) equal to the oscillating piston speed, which, in the absence of attenuation effects, is also the magnitude of the oscillating gas speed, U_g. That is, $\bar{v}_{gz}(z = 0) = U_g \sin\omega_a t$, where $\bar{v}_{gz}$ represents the z component of the gas speed averaged over velocity space. Although the Wang-Chang and Uhlenbeck distriubtion function for the gas is limited to small amplitude oscillations, the function does appear to have the correct structure since the moments of the distribution function yield reasonable expressions for the density, pressure, and temperature of the gas. The structure of f_g is important to this development of determining f_i since the effect of the sound wave on the suspended particles is through collisions of the particles with the gas molecules; that is, through the collision integral in the Boltzmann equation for the i^{th} particles. (This indirect coupling occurs because the mass loading for the particles has been taken to be very low.) As in the electric field case, the particle-gas collisions are modeled with an inverse fifth power force law. Again, using an approximation technique with a three term expansion to solve the Boltzmann equation, the particle distribution function is

$$f_i \simeq D_i\left\{\left[\alpha + \frac{c}{A^2}v_{iz} + \delta v_{iz}^2\right]\left[\exp(\zeta\cos\theta)\right]\left[\cos(\sigma\cos\theta - \chi)\right]\right\}(f_i)_M,$$

where

$$\alpha = 1 - \frac{c^2 - A^2}{2A^2} + \frac{2C_1 c[\zeta + \sigma\tan(\sigma\cos\theta - \chi)]}{\omega(\zeta^2 + \sigma^2)} \; ,$$

$$\delta = \frac{c^2 - A^2}{2A^2} - \frac{2C_1 c[\zeta + \sigma\tan(\sigma\cos\theta - \chi)]}{\omega A^2(\zeta^2 + \sigma^2)} \; ,$$

$$\zeta = \frac{C_2 A^2}{\omega_a c}\left[\frac{6c^2 - \frac{25}{3}A^2}{c^2 - 3A^2}\right],$$

$$\sigma = -\frac{C_2 A^2}{\omega_a(c^2 - 3A^2)}\left[40(c^2 - 3A^2) - \frac{1}{c^2}\left(6c^2 - \frac{25}{3}A^2\right)^{1/2}\right], \text{ and}$$

$$C_2 = \frac{C_1 U_g}{A^2}$$

In these equations $\theta = \omega_g(t - z/c)$, $A^2 = kT/m_i$, D_1 is an arbitrary constant, and as before $(f_i)_M$ is the Maxwellian function for the i^{th} aerosol species. A_2 is another collisional constant for the particle-gas collisions and equals 0.436 and D_1 is an arbitrary constant, as is the phase angle χ. The other quantities were identified earlier. As before, the expression for f_j is identical to f_i with a change of subscripts from i to j.

One property of this particle distribution function is that it exhibits a drift phenomenon due to the $\exp(\zeta\cos\theta)$ term. This stems from the $\sin\theta$ factor in the gas distribution function, which reflects itself through the collision integral of the Boltzmann equation. This factor subsequently becomes a coefficient in the set of partial differential equations that are solved for the expansion coefficients of f_i.

As before, the collision rate is determined by evaluating the collision integral for i-j particle collisions. We again assume that each species is charged. Using the hard sphere Debye-Hückel shielded potential model to account for the interaction of the charged particles during collisions, the collision rate, L_A, for the case where the two species are the same size (i.e., $R_i = R_j$) and have the same, but opposite average charge per particle is

$$L_A = D_1^2(0.23776 \frac{c^2}{A^2} - 0.99420) \; [\exp(2\zeta\cos\theta)][\cos^2(\sigma\cos\theta)] \; L_T,$$

where all of the terms have been introduced before. (This collision rate may also be used for uncharged, hard sphere particles by setting $q_i = 0$ in the charged particle interaction parameter, Ω.)

D. The Behavior of L_E/L_T and L_A/L_T

A plot of the maximum ratio of the self-collision rates, L_E/L_T and L_A/L_T, is given in Figures 1 and 2 for $R_i = 10^{-2}$ μm sized particles. The arbitrary constant, D_o^2, in the electric field case is eliminated by setting the ratio L_E/L_T to unity at zero electric field. The constant, D_1^2, in the acoustic case is eliminated by normalizing all the values of the ratio L_A/L_T with the value of the ratio evaluated for $U_g = 0$, which is taken to correspond to the state where collisions result from thermal motion only. The maximum values are plotted, instead of the time averaged values, because in both cases all points in the suspension will experience the total fluctuations of each field.

The curves in Figure 1 indicate that at a frequency of 10^3Hz the self-collision rates increase when the magnitude of the electric field in-

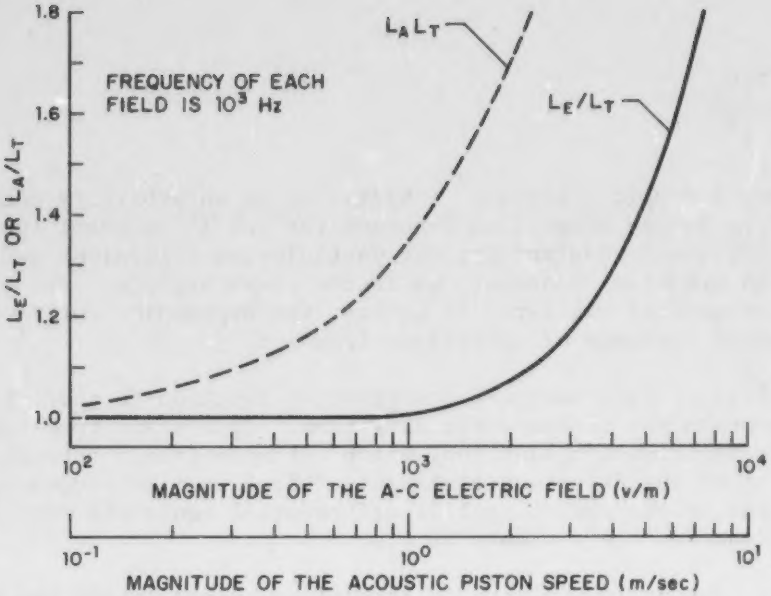


FIG. (1) SELF-COLLISION RATES AS A FUNCTION OF FIELD MAGNITUDES FOR $R_i = 10^{-2}\ \mu m$ PARTICLES

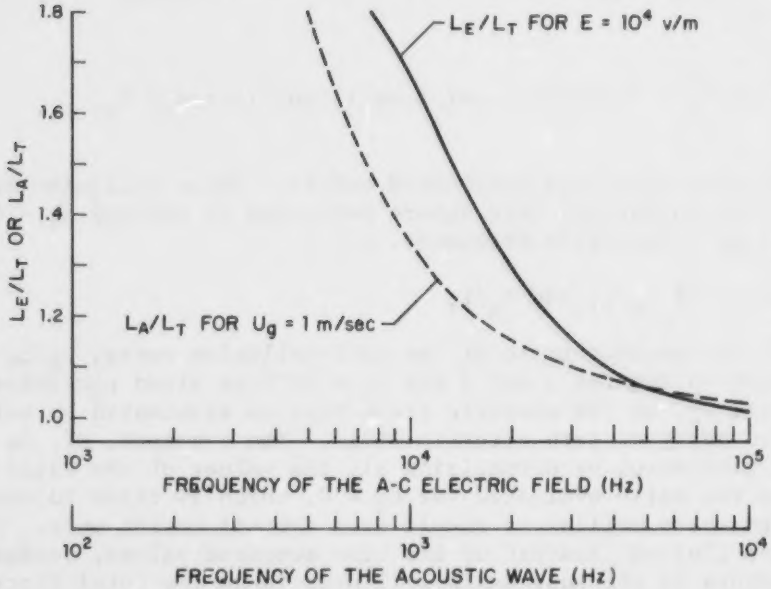


FIG. (2) SELF-COLLISION RATES AS A FUNCTION OF FREQUENCY FOR $R_i = 10^{-2}\ \mu m$ PARTICLES creases above about 10^3V/m or when the piston speed increases above about 0.5 m/sec. At room temperature the mean thermal speed of 0.01 μm sized particles is about 1 m/sec and, for a particle charge of one electron, the maximum speed of the particles in an electric field of strength 10^3V/m is about 6 m/sec. Then, as expected, the values of the collision rates increase above the thermal rates only when the directed motion due to the external fields is on the order of, or larger than the mean thermal speed of the particles. The curves in Figure 2 show that as the field frequencies increase, the ratios of the collision rates decrease and approach unity. This effect is due to the inertial lag of the particles which increases as the frequency of oscillations increases.

At this point, the simple kinetic model based on a dilute model of hard sphere particles suspended in a gas with the two external fields seems to be qualitatively correct. (The collision rates for the case of unlike particle collisions (that is, for $i \neq j$) are currently being computed.) The next step is to construct a similar model and include both fields simultaneously. In the meantime, the rates discussed above are going to be quantitatively checked during an experimental investigation involving self-collisions of monodisperse aerosols. The experimental program is discussed next.

III. EXPERIMENTAL INVESTIGATION

A. Introduction

The objective of the experimental investigation is to determine the collision rate and growth characteristics of submicron airborne suspensions subjected to an a.c. electric field, to an acoustic field, and to a combination of these two external fields. During the study, liquid and solid particles are to be used and the following parameters are to be varied: particle composition, concentration, and size; average electrical charge per particle; and the magnitude and frequency of each field. Monodisperse aerosols are to be studied primarily, but some simple polydisperse aerosols are to be examined later in the project. At the present time no coagulation studies have been carried out in the system. However, the experimental system is nearly ready for some initial self-collision rate measurements. The experimental system is described in the next section, followed by a discussion of the velocity and turbulence measurements recently completed. The last section describes the experimental program.

B. Experimental System

The experiments are to be carried out in a continuous flow coagulation system which is shown schematically in Figure 3. The aerosol is to be generated by a combination of an atomizer, particle conditioner, and an electric mobility classifier. This subsystem is capable of producing highly monodisperse particles in the size range from about 0.01 to 0.3 μm. The aerosol generating subsystem is discussed in more detail in a later paragraph. The generated aerosol is mixed with filtered air in the mixing chamber to increase the volume flow rate. From the mixing chamber the aerosol enters the

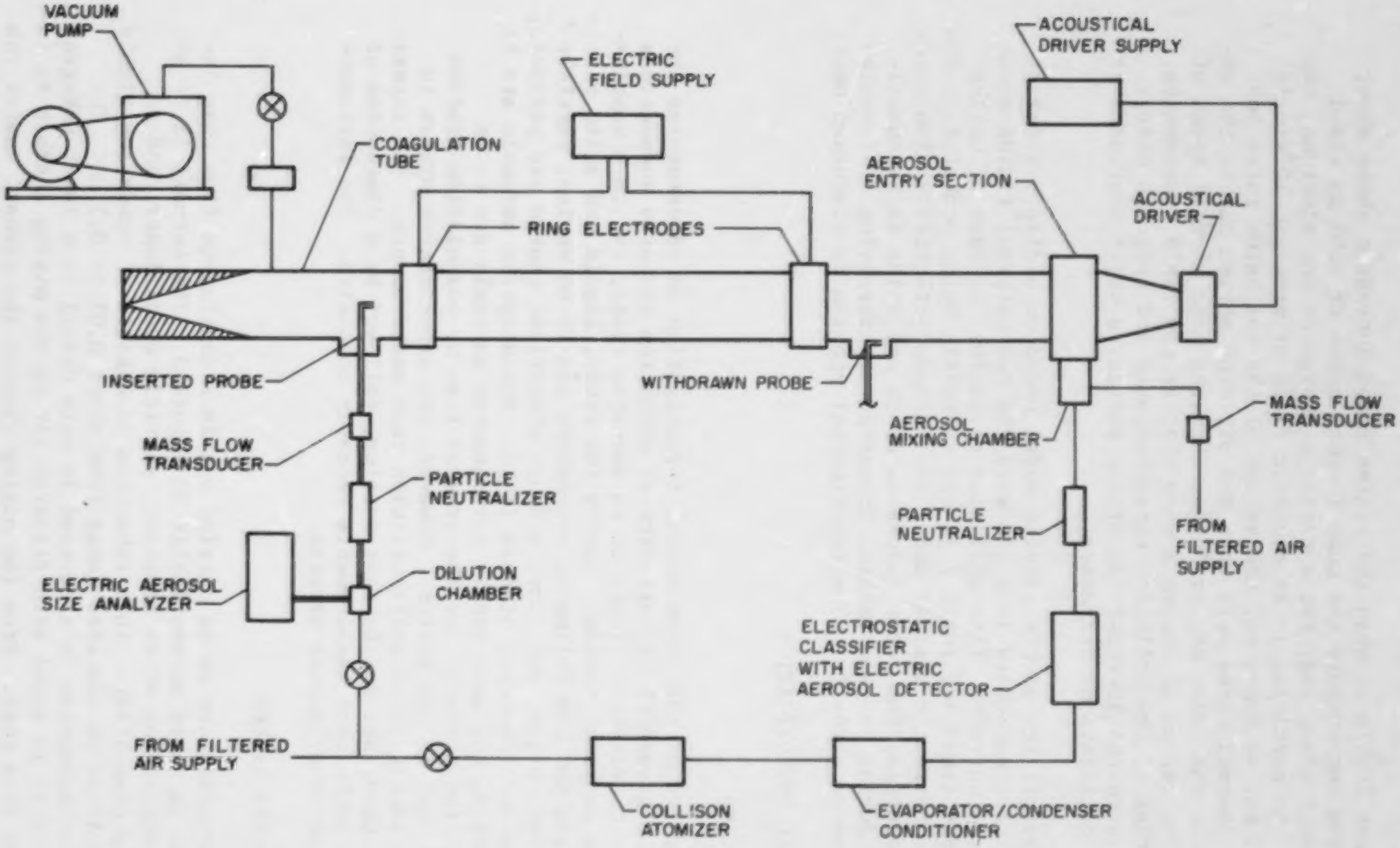

VACUUM PUMP
ELECTRIC FIELD SUPPLY
ACOUSTICAL DRIVER SUPPLY
COAGULATION TUBE
RING ELECTRODES
AEROSOL ENTRY SECTION
ACOUSTICAL DRIVER
INSERTED PROBE
WITHDRAWN PROBE
AEROSOL MIXING CHAMBER
MASS FLOW TRANSDUCER
MASS FLOW TRANSDUCER
PARTICLE NEUTRALIZER
PARTICLE NEUTRALIZER
FROM FILTERED AIR SUPPLY
ELECTRIC AEROSOL SIZE ANALYZER
DILUTION CHAMBER
ELECTROSTATIC CLASSIFIER WITH ELECTRIC AEROSOL DETECTOR
FROM FILTERED AIR SUPPLY
COLLISON ATOMIZER
EVAPORATOR/CONDENSER CONDITIONER

FIG. (3) EXPERIMENTAL SCHEMATIC OF THE CONTINUOUS FLOW COAGULATION SYSTEM.

horizontal coagulation tube through an inlet section. The continuous flow coagulation tube is shown in more detail in Figure 4. The coagulation tube is about 4.9 m long and is made of 6 inch (inside diameter) Plexiglas tubing. The tube has two ports for isokinetic sampling probes, two ring electrodes wrapped around the outside of the middle section, plus an electro-mechanical, acoustical driver mounted at the end with the aerosol inlet section. The acoustical driver is to generate a continuous wave of traveling sound waves which travel down the coagulation tube and which are absorbed by a fiberglass mat located at the far end. The acoustical driver is a James B. Lansing professional series compression driver, Model 2440, fitted with a specially machined, air tight, exponential horn which makes a smooth transition from the 2 inch diameter of the driver throat to the inside diameter of the coagulation tube. A thin membrane of plastic wrap is to be used to keep the aerosol particles out of the driver. The acoustical driver is to be driven by an audio power amplifier with an output up to 60 watts. A signal generator is to be used as the input to the amplifier. The ring electrodes, which are spaced about 150 cm apart, are used to generate the axial electric field. The a.c. electric power supply consists of an inhouse built high voltage amplifier (0 to 6 kV) driven by the output from a signal generator over the frequency range of about 30-400 Hz. Two moveable, isokinetic sampling probes are used (one at a time) to extract a continuous sample from various points across the diameter of the coagulation tube at the entrance and exit planes of the test section. The distance between the probes is about 183 cm, which is the length of the test section within the coagulation tube. The extracted sample is diluted with clean air and passed through an electric aerosol size analyzer to determine the particle size distribution. (More detail on the probe design and the analyzer is provided below.) The aerosol is continuously pumped through the coagulation tube and the pressure can be varied about ± 5 inches of water using a vacuum pump. As mentioned earlier, this feature of a continuous flow minimizes transient effects that have been known to cause erroneous results in past studies that depended on time decay measurements to determine collision rates.

The coagulation tube is a critical component for it is important that the flow in the tube be uniform and relatively free of turbulence. As shown in Figure 4, the flow enters the tube through the entry section which introduces the flow uniformly and axially so as to minimize turbulent effects. The distance from the entry section to the inlet of the test section (i.e., the location of the upstream isokinetic probe) is 60.96 cm and the distance from the inlet to the outlet of the test section is 182.9 cm. For a flow rate of 44 lpm and a tube diameter of 15.24 cm, the Reynolds number is about 430. Assuming that the flow is totally uniform at the entrance of the coagulation tube, the thickness of the laminar boundary layer at the test section inlet is estimated to be about 1.01 cm and at the outlet the thickness is 4.04 cm. Therefore, the flow is not fully developed in the test section and there should be a central core of nearly uniform axial flow with a uniform speed slightly greater than the average speed given by the flow rate divided by the cross-sectional area of the tube. (Velocity and turbulence measurements have been taken across the inlet and outlet cross-sections using a hot film anemometer inserted through the sampling tube ports and the results are to be discussed in a later paragraph.)

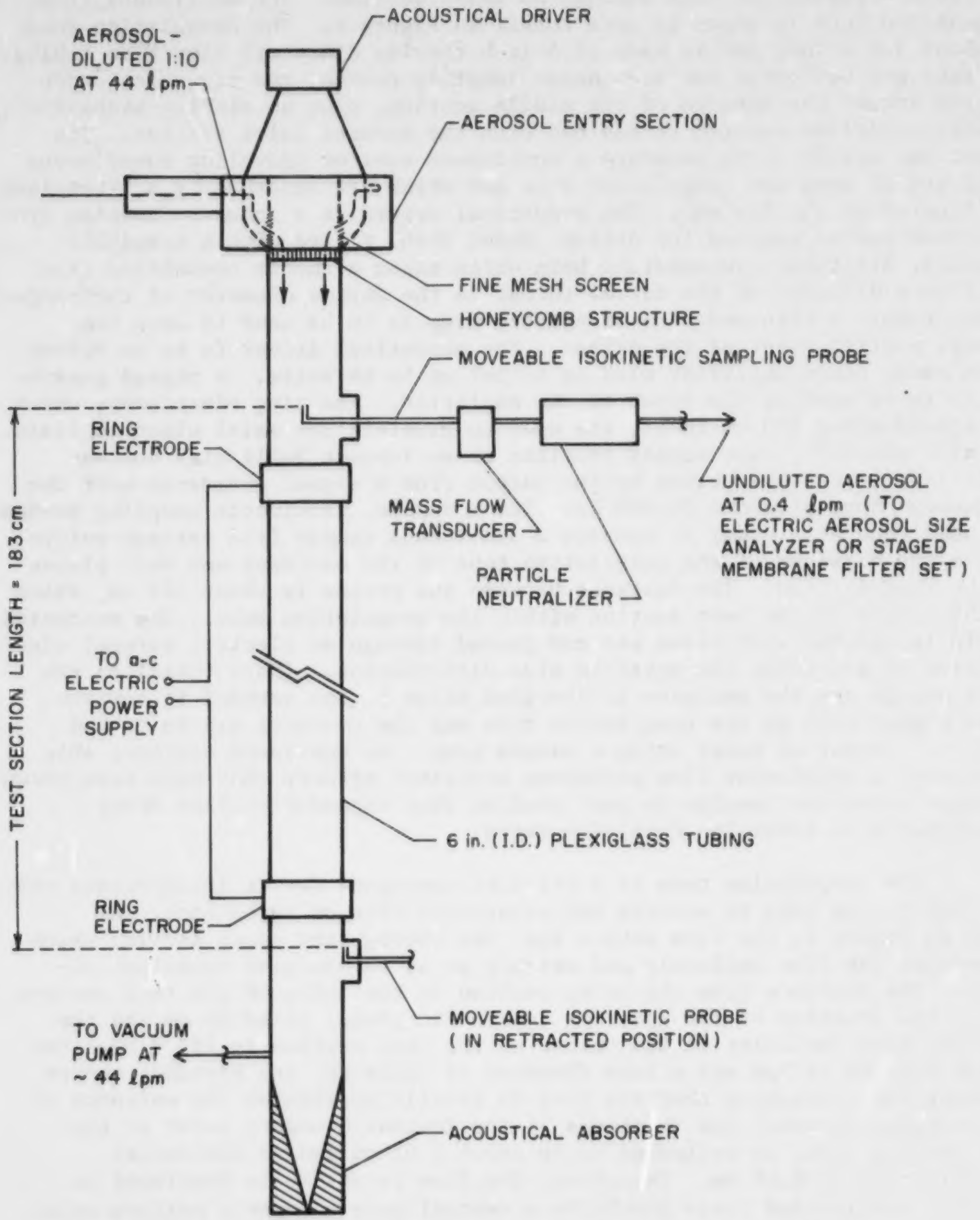

ACOUSTICAL DRIVER
AEROSOL-DILUTED 1:10 AT 44 ℓpm
AEROSOL ENTRY SECTION
FINE MESH SCREEN
HONEYCOMB STRUCTURE
MOVEABLE ISOKINETIC SAMPLING PROBE
RING ELECTRODE
UNDILUTED AEROSOL AT 0.4 ℓpm (TO ELECTRIC AEROSOL SIZE ANALYZER OR STAGED MEMBRANE FILTER SET)
MASS FLOW TRANSDUCER
PARTICLE NEUTRALIZER
TEST SECTION LENGTH = 183 cm
TO a-c ELECTRIC POWER SUPPLY
6 in. (I.D.) PLEXIGLASS TUBING
RING ELECTRODE
MOVEABLE ISOKINETIC PROBE (IN RETRACTED POSITION)
TO VACUUM PUMP AT ~ 44 ℓpm
ACOUSTICAL ABSORBER

FIG. (4) SCHEMATIC OF ENTRY SECTION AND COAGULATION TUBE.

The aerosol is to be sampled and analyzed at different positions across the coagulation tube diameter at the inlet and outlet stations of the test section. The samples are to be taken with the use of isokinetic probes (inserted one at a time so as to minimize flow disturbances). The diameter of the probes is 1.45 cm which gives a probe flow rate of 0.4 lpm at the average flow velocity in the tube. A dilution system (shown in Figure 3) is used to increase the total flow rate to the analyzer to a value of 4 lpm. This corresponds to a concentration dilution of about 1:10.

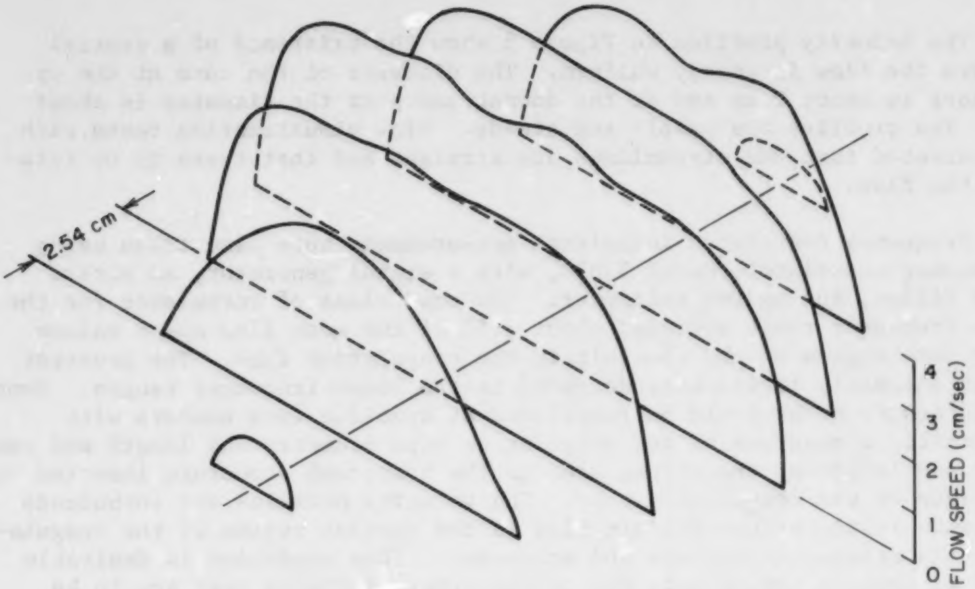
The aerosol generating equipment is commercially available from Thermo-Systems Incorporated of St. Paul, Minnesota, and is based on the submicron aerosol generating system developed by Liu and associates at the University of Minnesota. (The system is described and discussed by Liu and Pui in Reference [10].) The system consists of an atomizer, an evaporator-condensor conditioner, and electrostatic classifier with an electric aerosol detector, and a neutralizer. The atomizer can generate a moderately monodisperse aerosol from non-volatile liquids (such as DOP) dissolved in ethyl alcohol and from aqueous solutions such as methylene blue. Suspensions of pre-sized latex spheres and of carbon (non-spherical) and iron oxide (sperical) particles with a mean particle size in the range from 0.01 to about 0.3 μm may also be produced by mixing these particles in very pure demineralized water, or in alcohol. (All of these particles are to be studied during the coagulation tests to be described later.) The evaporator-condenser is included to convert the moderately monodisperse aerosols produced by the atomizer from liquids (like DOP) into a more monodisperse suspension by an evaporation process followed by a condensation process. (The evaporator-condenser is not to be used for the solid particle suspensions.) The electrostatic classifier serves as the final stage to select particles, on the basis of their electric mobility, to make up a truly monodisperse aerosol. An electric aerosol detector is included so that the particle concentration of the generated monodisperse aerosol may be determined when desired. (As discussed later, the concentration measurements are required in order to calibrate the electric aerosol analyzer used with the isokinetic sampling probes.) A Kr-85 neutralizer is also shown in Figure 3 and is to be used when a neutral aerosol is desired for the coagulation tests. Liu and Pui [10] indicate that this system is capable of producing uniform particles from 0.01 to 0.5 μm at concentrations up to 10^6 cm^{-3}. (The manufacturer indicates that concentrations up to 10^7 are possible with DOP aerosols.) According to Liu and Piu, the particle diameter of the aerosols generated by this system may be determined to within an accuracy of about 2% and the concentration to within about 5%. The relative standard deviation of the size distribution for aerosols produced by this system ranges from about 0.04 to 0.08.

The particle size analyzer is an Electric Aerosol Size Analyzer (EAA), Model 3030, manufactured by Thermo - Systems Incorporated and is an instrument developed by Whitby and Liu at the University of Minnesota. The analyzer first electrically charges the incoming aerosol particles and then sizes them on the basis of their electrical mobility in the presence of an electrostatic field. The performance of the analyzer is presented in a paper by Liu and Pui [11]. The analyzer has been purchased with two additional voltage divider chips: one to divide the 0.02 to 0.22 μm size range into ten arithmetic size intervals of 0.02 μm each, and one to divide the 0.20 to 0.40 μm size range into ten arithmetic size intervals of 0.02 μm each. (The standard voltage divider chip divides the 0.0032 to 1.0 μm size range into ten geometrical size intervals that vary in size from 0.0024 μm to 0.438 μm.) This modification is included in order to improve the size resolution of the instrument. Using the aerosol generating system described above, the EAA is to be calibrated by connecting the analyzer directly to the output of the aerosol generator. It is expected that through a careful calibration procedure, the size analyzer may be used to measure unknown concentrations in the individual channels to within an accuracy of about 5%, which is the concentration accuracy of the submicron aerosol generating equipment discussed above.

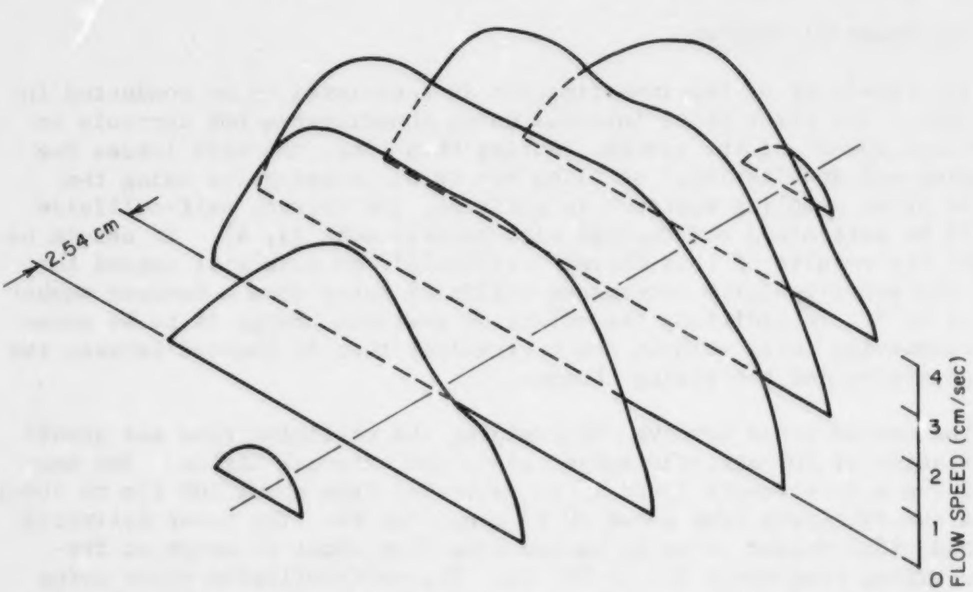
A staged membrane filter set is also to be used in conjunction with the isokinetic sampling probe to extract and collect particle samples for analysis of particle shape, size uniformity, and cluster configuration. The membrane filters are to be arranged in order of decreasing pore size from 0.8 to 0.025 μm. Photomicrographs of the collected samples are to be used to check the EAA results and to determine cluster configuration during the coagulation studies using solid particles. (In addition, an attempt will be made to borrow a portable diffusion battery and a condensation nuclei counter during the investigation to provide an additional check on the EAA and on the filter- photomicrograph data.)

C. Velocity Profiles and Turbulence in the Coagulation Tube

A hot film anemometer has been used to analyze the flow through the coagulation tube. Filtered air without any particle matter was used during the tests. The turbulence structure was initially unsatisfactory and the flow was not uniform. Following the addition of a screen in the inlet section and a honeycomb structure across the coagulation tube at the inlet section, the flow was found to be uniform and the turbulence fairly low. Velocity profiles at the upstream and downstream locations of the isokinetic probes (i.e., at each end of the test section) are shown in Figure 5. The profiles were taken using a probe mounted on a 90° support sting that could be rotated and translated in order to survey the cross-sectional area of the coagulation tube at each location. The probe sting passed through the isokinetic probe port and "O" rings were used to ensure that there was no leakage from the coagulation tube. The probe used for the measurements was a Disa fiber film probe, Model 55R11. The hot film was operated in the constant temperature mode by a Disa anemometer, Model 55D01. Using the square wave technique, the frequency response of the hot film anemometer was about 11 kHz at a mean flow speed of 4 cm/sec. The anemometer output was not linearized and the system was calibrated using the fully developed flow in a 1.50 cm diameter tube with an L/D ratio of 80. The flow rate through the calibration tube was measured using a Sierra Instruments mass flow transducer, Model 500-3, which was calibrated to within an accuracy of 1%. A signal of 510 mV corresponded to a speed of 4 cm/sec. It was found that the system could not resolve speeds less than about 0.8 cm/sec. The calibration curve was repeatable to within 3% and the velocity profiles within 7%.



VELOCITY PROFILES AT UPSTREAM PROBE PORT



VELOCITY PROFILES AT DOWNSTREAM PROBE PORT

FIG. (5) VELOCITY PROFILES AT UPSTREAM AND DOWNSTREAM ISOKINETIC PROBE PORTS.

The velocity profiles in Figure 5 show the existence of a central core where the flow is nearly uniform. The diameter of the core at the upstream port is about 9 cm and at the downstream port the diameter is about 6.5 cm. The profiles are stable and steady. Flow visualization tests with smoke indicated that the streamlines are straight and that there is no rotation of the flow.

Frequency correlated turbulence measurements were also taken using a Disa analog correlator, Model 55D70, with a signal generator, an active low-pass filter, and an rms voltmeter. The rms values of turbulence for the 0-11 kHz frequency range averaged about 3.8% of the mean flow speed values over the core region of the flow within the coagulation tube. The greatest turbulent intensity levels were detected in the lower frequency ranges. Some of the intensity peaks could be correlated at specific wave numbers with characteristic dimensions of the coagulation tube diameter and length and the diameter and length of the straws used in the honeycomb structure inserted at the entrance of the coagulation tube. The velocity profiles and turbulence measurements indicate that the air flow in the central region of the coagulation tube is relatively uniform and quiescent. This condition is desirable in order to isolate the effects due to the external fields that are to be investigated later. In this case, the effects of velocity gradients and turbulence should be negligible.

D. Experimental Program

The remainder of the investigation is envisioned to be conducted in three phases. The first phase involves using monodisperse DOP aerosols to calibrate and check out the system. During this time, the wall losses due to diffusion and gravitational settling are to be investigated using the isokinetic probe sampling system. In addition, the thermal self-collision rate is to be determined and checked with earlier work [3, 4]. It should be noted that the results of this thermal self-collision work will extend the range of the experimentally determined collision rates from a Knudsen number of about 1 to 6. In addition, the effect of particle charge is to be examined by conducting tests without the nuetralizer that is located between the aerosol generator and the mixing chamber.

The second phase involves determining the collision rate and growth characteristics of DOP aerosols subjected to the external fields. The magnitude of the a.c. electric field is to be varied from about 100 V/m to about 4kV/m and the frequency from about 30 to about 400 Hz. The power delivered to the acoustical driver is to be varied from 0 to about 60 watts at frequencies ranging from about 300 to 10^4 Hz. The self-collision rates using a single monodisperse DOP aerosol are to be determined for each field and for the two fields together. It should be noted that regardless of which field is applied, the number of collisions per unit time between particles of the same size is expected to exceed the number due to thermal motion alone. This increase is due to the superposition of the thermal motion and the motion induced by either field. For the very fine particles (i.e., particles with radii less than 0.1 μm) the thermal motion is relatively large. When both fields are applied simultaneously, the number of collisions per unit time between particles of the same size exceeds the thermal rate because of the differential motion induced by both fields. Of course, the thermal motion will also contribute to the collision rate for the smaller particles in this case. In addition to the use of mohodisperse suspensions, some simple bimodal suspensions are also to be generated and studied. During these tests the charge on the particles is to be varied by removing the neutralizer located downstream of the aerosol generating subsystem and by inserting a corona charging stage in the aerosol mixing chamber. In all of these tests, photomicrographs of particle samples are to be used to check the EAA results for size uniformity. The samples are to be withdrawn through the isokinetic probes and collected by the staged membrane filter set.

The third phase involves repeating the second phase for different types of monodisperse particles. The different materials include latex, NaCl, carbon, titanium oxide, and iron oxide. Photomicrography of samples withdrawn through the probes is to be used to monitor particle size and shape, including cluster configurations in the case of the solid particle tests.

IV. SUMMARY

An experimental investigation has been described that involves the use of an a.c. electric field and a traveling wave sound field to increase the collision rate between particles in an aerocolloidal suspension. In addition, a simple free molecule regime theory for the collision rates due to each field applied separately has been discussed. Experimentally determined velocity profiles and average levels of turbulence in the continuous flow coagulation tube have also been reported. The plan for completing the investigation has been outlined.

The investigation is a basic study of the collision rates between particles with radii less than about 0.3 μm suspended in air and subjected to electric and acoustic fields. The results of the study should contribute to a better understanding of particle collisions and growth characteristics under the influence of electric and acoustic fields. In addition, it is anticipated that the results will show that these fields can provide an effective means for preconditioning man-made particulate suspensions so that the very fine particles may be removed in a subsequent and conventional particle removal device.

ACKNOWLEDGMENTS

The author is appreciative of the work of P. H. Paul who made the hot film anemometer measurements and L. W. Byrd who designed the inlet section for the coagulation tube. This study is supported by a research grant from the Solid and Particulate Processing Program of the National Science Foundation.

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WRITTEN DISCUSSION

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The experimental research efforts proposed by Scholz should permit the achievement of a new level of refinement in data quality. With reference to Scholz's Figure 4, the apparatus which is to be employed is as good as the state-of-the-art has to offer and the flow schematic seems to be well thought out. It may be anticipated that a high quality aerosol will result through uniformization of a Collision Atomizer spray with both an evaporator/

condenser and an electrostatic classifier. It is planned to use a specially equipped electrostatic aerosol analyzer which is fitted to divide the range of 0.2 - 0.4 micrometers into ten arithmetic intervals, however this resolution may demand more of the instrument than it is capable of providing.

With respect to the analytical modeling which is presented, p. 4, the effort to date has been directed towards the free molecular regime with separate consideration of the electric and acoustic fields. The proposed next step, that of modeling the effects of the combined fields, should provide an interesting extension to the understanding of aerosol technology.

The predictions which have been made of the augmentation to thermal collision rates effected by acoustic and electric effects (Figures 1 and 2) show that for particles of 10^{-2} micrometers radius the expectation is that the rates would be approximately doubled -- at least for the operational conditions which were examined. Considering that it is necessary to have on the order of 10^6 collisions to cause a 10^{-2} micrometer droplet growth to one micrometer, a substantial augmentation is needed.

OPEN DISCUSSION

Scholz: I have a couple of comments to make regarding the taxation of the electrostatic aerosol size analyses. I cannot say at this time what the detection range would be. I had an impression that the system is capable of producing aerosols with concentration up to 10^6 particles per cubic centimeter. I do not know how good that value is in terms of different sizes and different materials. At this point all I can say is that we are expecting to get good results.

For the comment on the particle size under 100 μm not being terribly important; I would like to indicate that I used the 100 μm particle size as a demonstration to show the significance of the thermal motion. Our approach is to try to understand the interaction of forces with the particles and to try to get a good handle on the basic mechanism taking place within the aerosol. It is from this point of view that we used the 100 μm size particle as a sample.

Baril: We have some work going on in our laboratory which is comparable to your study. I was wondering if you are familiar with Professor Massudu's work on hammering particles with traveling waves. I gather that you are going to use this electric field to enhance the traveling wave. We have done extensive work in this area. If you would increase the number of rings you could have a much better effect. In fact you can transport the particles down the tube without the acoustical energy. You might even be able to take the air energy out and just use the electric field. If you change the size of the rings as you go down, it will make the particles come to the center of the system.

Scholz: I am not familiar with any of this. However, I am very much interested in learning more about your research in this area.

Helfritch: Paul, is the inverse force law the right one to use between a charged particle and a non-charged gas molecule?

Scholz: I do not exactly know the answer. In kinetic theory, the inverse power force law is the usual one and seems to give qualitative effects, if you are interested in qualitative interaction. I used it basically as a standard model. I do not know how bad it is in terms of a neutral versus a charged particle interactional whether it is second order or primary, I do not know.

Reif: Back in the 1940's, Pohl did quite a bit of work with sonic dispersion of fog. He reportedly had a system that worked on ships which would disperse fog up to 300 meters. Are you familiar with this work and if so, whatever became of it.

Scholz: I am not familiar with the work, but there may be some people here who are.

Baril: I might add something to that, Bob. There was also some work done at Orly airport by Dr. Bouche with sirens. If I remember correctly, this all fell by the wayside. It turned out that the amount of energy and money put into these systems was astronomical.

Shaw: I can comment on that. It was proven that if you use seeding techniques it is much more economical for that particular application than acoustic sirens. Some redesign work was done by Calspan Corporation.

Liu: I am a little mystified with the use of the Debye length for this calculation. It would seem to me that it is applicable primarily to the plasmas and colloidial suspensions where the bipolar ions are around the particles. When you are dealing with a gaseous medium, how does the Debye length come into play?

Scholz: It is my understanding that as long as the Debye length is greater than the mean distance between the particles, then the Debye particle theory may be applied. The other necessary condition is that the energy in the electrostatic field of the particle at one Debye length must be less than the thermal energy that the particle has. Both of these conditions are satisfied in our application.

In terms of polarization effects and so forth, I don't know. I am only saying that it is a model. In terms of kinetic theory, the coulomb interparticle force law has an infinite cross section. In order to treat the system you have to cut it off at some point. Kinetic theory is based on two-particle interaction. It is necessary to cut off the collision interaction length at some point. We used the Debye length.

Liu: The Debye length depends on the concentration of the charge carriers and the particles.

Scholz: The concentration, dielectric strength, the thermal characteristics are also important.

Hidy and Brock have a section in their book on the application of the Debye Potential Theory in aerosols and have used it.

Cooper: In the experimental set-up, the plexiglass, unless you are careful, is going to have areas of localized charge. When you apply the voltage on the rings, you will get stress charging which can create more problems. The use of isokinetic samplers can also cause problems since it is crucial at which position they should be isokinetic. The right angle bends are going to cause some difficulties. Precisely when the particle relaxation time times its velocity is comparable to the dimension of the sampler. Those bends look as though they are going to give substantial impaction on the inside surface.

Scholz: In reference to the plexiglass, at this time we have only used low field intensities, and we have not noticed any localized effects. It is possible that changing to higher intensity might cause some problems.

Cooper: You may notice localized areas of intensities because the fields may be as high as a thousand volts per centimeter.

ENGINEERING ASPECTS OF DRY MAGNETIC SEPARATION

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ABSTRACT

Engineering aspects of batch-type and continuously operating magnetic gas-particulate separation methods are compared. The comparison is made in conceptual applications to desulfurization of dry fine-pulverized pretreated coal. With pretreatment, it is claimed that magnetic separability of mineral sulfur forms occurring in coal can be considerably enhanced so that a variety of magnetic methods could be expected to be applicable. In processing fine pulverized coal, batch-operated HGMS-type separators offer the advantage of efficient desulfurization of high coal-value recovery but can incur practical difficulties associated with packed bed operation. By contrast, continuously operating smooth-bore magnetic separators have the advantages and simplicity of continuous operation but are not as effective in desulfurization in high gas-velocity applications. Trade-offs between separator performance and total process requirements will be very important in future developments of magnetic methods for gas-particulate separations. New applications of magnetism can be anticipated in process industries employing entrained and fluidized bed operations.

INTRODUCTION

Two widely different approaches have been considered for development in the application of magnetic methods to gas-particulate separations. Basically, these two approaches provide the researcher with a choice between batch-operated magnetic filtration (modified HGMS-type separators) or continuously operating smooth-bore beam-splitting-type devices (e.g., multipole magnetic separators). These magnetic separators can be expected to have widely different capabilities in gas-particle separations and to exhibit significantly different process design requirements in high velocity and high throughput applications to gas clean-up.

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The batch-type magnetic separator envisaged for gas-particulate separation is similar to the HGMS-type separator recently developed for wet minerals beneficiation. An example of the HGMS-type separator is illustrated in Figure 1[1]. This device utilizes flow through a relatively open and porous magnetized bed which is composed of filimentary strands of strongly magnetic material such as ferritic stainless steel wool, wire mesh, or expanded metal lath. The magnetized bed elements provide a volume distribution of very short range but intense magnetic capture forces which can be made specific to particle size. These batch-operated devices have proven commercially effective in wet separations of feebly paramagnetic minerals with size distributions occurring into the submicron range [2].

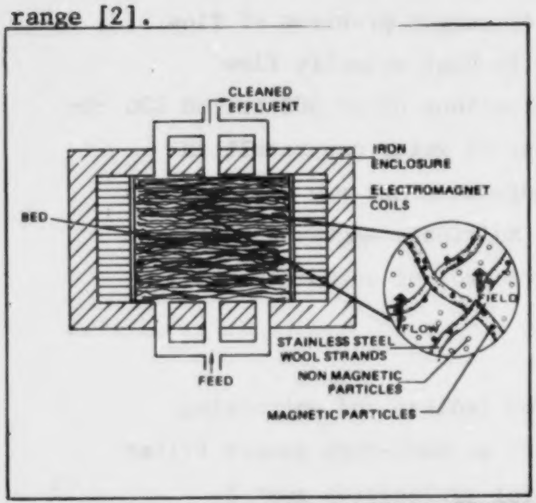

CLEANED
EFFLUENT
IRON
ENCLOSURE
ELECTROMAGNET
COILS
BED
FIELD
FLOW
STAINLESS STEEL
WOOL STRANDS
FEED
NON MAGNETIC
PARTICLES
MAGNETIC PARTICLES

Figure 1. VERTICAL SECTION THROUGH CENTER OF HIGH GRADIENT MAGNETIC SEPARATOR

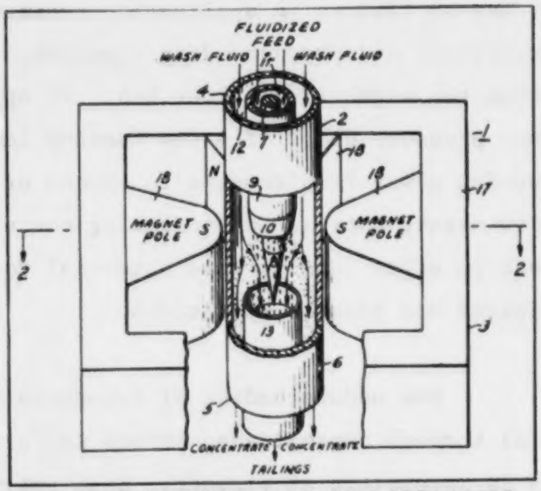

FLUIDIZED
FEED
WASH FLUID
WASH FLUID
11
4
2
1
12
7
18
N
18
18
17
9
MAGNET
POLE
S
10
S
MAGNET
POLE
2
2
13
3
6
5
CONCENTRATE
CONCENTRATE
TAILINGS

Figure 2. QUADRUPLE MAGNETIC SEPARATOR (Ref. [3])

The second approach to dry magnetic separation of gas-entrained particulates employs continuously operating smooth-bore magnetic beam-splitting devices utilizing multipole magnets of the type presently employed in particle physics applications [3,4]. These devices, which have not yet been applied outside of the laboratory, offer the potential simplicity of continuous smooth-bore operation. The magnetic forces they generate, however, are weaker and less specific to fine-size materials and they are inherently power-intensive devices. An example of a quadrupole magnetic separator is shown in Figure 2[3].

HGMS-type structures are probably closer to commercial development in dry handling of fine pulverized materials than are continuously operated magnetic devices. The latter type separators, however, are deserving of consideration because of several significant technical reasons. They require no gas flow interruption, are relatively immune to abrasive wear, and are low-pressure drop devices. In gas-particle transport, entraining velocities can be expected to be very high because of the large density difference between the suspended particles and the fluidizing gas. Even though modified HGMS-type devices are superior to multipole structures from the standpoint of fine particulates removal, they may be limited in application because of severe problems of flow distribution and pressure drop associated with high velocity flow through the magnetized filter bed. In applications of 50 000 to 100 000 cfm flows, pressure drops of a few hundred inches of water can result in drive-fan power requirements in excess of megawatts and may well overshadow magnet capital and operating costs. Multipole magnet structures appear to offer considerable potential for control of entrained and fluidized bed reactor operations.

The understanding of two-phase flow (solids and entraining fluid) through consolidated porous media such as HGMS-type packed filter beds is incomplete at present. Each potential application must be treated individually with due consideration for the magnetic particle density, mean size, and size distribution and for the size and shape of individual magnetic filter bed elements. Preliminary engineering evaluations indicate that quadrupole-type smooth-bore magnetic separators may possibly compete with batch-operated filter-type magnetic separators in high velocity separations of strongly magnetic micron-size materials. To be competitive in separations applications, however, it is believed that multipole structures will have to be developed with pole-face field strengths of the order of 100 000 Oe at one foot pole-opening.

A comparison of these magnetic separation methods will be given in a conceptual application to the removal of dry mineral sulfur from fine-pulverized coal.

MAGNETIC DESULFURIZATION OF DRY COAL

Coal has traditionally been cleaned to control mineral matter content. Cleaning for sulfur control is now becoming increasingly important because of governmental regulations of sulfur oxide emission in coal combustion flue gases.

Sulfur occurs in coal both as a discrete mineral form (predominately iron disulfide [FeS_2]) and as "organic sulfur" which is believed to be chemically bound to the coal structure. Because of the need to remove both sulfur forms, chemical methods are being developed to augment more conventional physical cleaning methods which are generally restricted to removal of coarse (>35 mesh) mineral sulfur only. Recent studies have shown, however, that chemical cleaning per se can be expected to be expensive [5]. For this reason, there is renewed emphasis on upgrading state-of-the-art physical cleaning technologies and on developing new approaches to combined physical and chemical cleaning which have the capability of preparing a moderate cost clean fuel without requiring coal conversion.

Coal washability studies [6] have shown that mineral sulfur occurs in U.S. coals in a great variety of particle sizes. Most of the important Eastern utility grade coals have significant amounts of mineral sulfur with particle sizes occurring down into the micron range. Any comprehensive new coal cleaning method developed for these coals, then, will of necessity be required to process coals which have been pulverized to sizes generally finer than 100 mesh (75 microns). There is no such commercial coal cleaning method in use today.

Laboratory investigations have demonstrated the scientific and technical feasibility for magnetic desulfurization of coal [7]. Those investigations show that the mineral sulfur occurring in coal is paramagnetic and is magnetically separable from mineral-free coal which is diamagnetic. That work was not pursued, however, because the laboratory magnetic separators employed could not be adapted economically to commercial-size applications. With the recent development of HGMS-type devices, however, there is now a magnetic separation method available for adaptation to desulfurization of coal with the potential for economical removal of micron-size mineral contaminants in commercial-scale installations.

Because of this potential, the concept of magnetic desulfurization of coal is again being pursued in several laboratory investigations. The results reported here were derived in an engineering evaluation of dry magnetic methods for desulfurization of pneumatically conveyed fine-pulverized coals.

The most direct use of magnetic methods in dry coal desulfurization would be in power plants which burn fine-pulverized coal. The magnetic separator would be "engineered" to fit in line between the pulverizer and the coal burner inlet. A preliminary evaluation of this concept [8], however, has indicated that the use of magnetic separators in this application may not be practicable because of severe limitations imposed by the feeble paramagnetism of the mineral sulfur and the use of very high carrier gas velocities (5 000 to 6 000 ft/min). Estimates indicate that capital and operating costs for this direct approach to magnetic desulfurization may be similar to those of stack gas scrubbing. These costs are largely associated with the high magnetic field strength required to capture paramagnetic minerals and with power requirements for auxiliary equipment. Because of this unfavorable cost estimate, the engineering evaluation presented here was restricted to the use of magnetic separators in the processing of dry coals which have been pretreated to make the mineral matter strongly magnetic.

MAGNEX PROCESS

Several methods have been proposed for converting mineral matter in coal to strongly magnetic forms [9]. One new method is the Magnex process which exposes dry pulverized coal to gaseous iron pentacrbonyl [$Fe(CO)_5$] under conditions whereby the impurity mineral matter becomes strongly magnetic and can be separated from the coal by magnetic separation methods [10].

Under the Magnex reaction conditions, gaseous iron pentacarbonyl is substantially decomposed into free iron and gaseous carbon monoxide. It is claimed that the liberated iron reacts with the surface of the pyrite mineral to form a strongly ferrimagnetic pyrrhotite-like sulfurous mineral. The reaction stoichiometry is assumed to be:

$$\underset{\substack{\text{iron}\\\text{disulfide}\\\text{(pyrite)}}}{FeS_2} \quad + \quad \underset{\substack{\text{iron}\\\text{pentacarbonyl}}}{xFe(CO)_5} \xrightarrow{170\text{-}200^\circ C} \underset{\text{"iron-rich disulfide"}}{[(1-y)FeS_2] \cdot [(y+x)FeS_{\frac{2y}{x+y}}]} + \underset{\substack{\text{free}\\\text{carbon}\\\text{monoxide}}}{5xCO} \qquad (1)$$

If x moles of iron pentacarbonyl react with 1 mole of granular and completely liberated iron pyrite, it is assumed that a fraction of y moles of the outside of the pyrite grain react to give y+x moles of an iron-sulfide of sulfur to iron atomic ratio of 2y/(y+x). If one demands that this ratio correspond to that of magnetic monoclinic pyrrhotite ($\sim Fe_7S_8$), then the mole fraction y is determined by the mole quantity of iron pentacarbonyl used, x (see iron-sulfide phase diagram[11]). This relationship is shown in Table I where percent conversion of iron pyrite to monoclinic pyrrhotite is shown for a coal containing about 1.2 wt.% pyritic sulfur and where, for simplicity, it has been assumed that free iron is deposited only on the surface of the iron pyrites.

In the magnex process, it is conjectured that some of the free iron also coats the non-sulfurous mineral matter, making it amenable to magnetic separation as well. For this case, the conversions shown in Table I are upper limits expected for reactions with pyrites only. It is claimed that free iron does not adhere to the mineral-free coal structure.

Laboratory quality monoclinic pyrrhotite (Fe_7S_8) is strongly ferrimagnetic with a spontaneous magnetic moment of 13 emu/gm at 200°C which saturates at ≈4 kOe [1]. This moment is about 14 000 times greater than the paramagnetic moment of pure pyrite at 200°C in a 4 kOe magnetic field. Indeed, only about 0.7% of the pyrite need be converted for the magnetic moment of the "iron-rich disulfide" to predominate that of the converted mineral. The resulting impurity saturation magnetic moment calculated for assumed selected reaction of iron pentacarbonyl with iron-disulfide is shown in Table I. Magnex has recommended 32 lb iron pentacarbonyl per ton coal for most Eastern coals, so that with this concentration, one can, under ideal conditions, anticipate a magnetic moment of 8 emu/gm which saturates in a 4 kOe magnetic field.

TABLE I

"Ideal" Pyrite Conversion by Iron-Pentacarbonyl
for a Coal Containing 1.2 wt.% Pyritic Sulfur -- Dry Basis

Iron Pentacarbonyl Concentration lb $Fe(CO)_5$/T dry coal	Calculated % Conversion $FeS_2 \rightarrow Fe_7S_8$	Calculated Magnetic Moment (emu/gm) @200°C
0	0	0
20	36	5
32	59	8
40	73	9
55	100	13

IRON CARBONYL PROCESS

A process flow diagram for an iron-pentacarbonyl process is shown in Figure 3. This layout is similar to the Magnex process [10] except that here fine-pulverized coal is specifically employed and iron pentacarbonyl is consumed. In this process layout, carbon monoxide liberated in the reaction is flared to produce carbon dioxide which, in turn, is used for penumatic transport purposes. In a commercial process, the liberated carbon monoxide might be passed through an iron reservoir for generation of process iron pentacarbonyl. In this manner, iron rather than iron pentacarbonyl would be consumed. It is to be emphasized that the hypothetical process described here differs from the Magnex process primarily in that the coal is fine-pulverized. When coarser coals are treated with the Magnex process, pneumatic transport is not necessary and simple magnetic separators adaptable from other areas of minerals beneficiation can be employed [13]. The hypothetical process given here for desulfurizing fine-pulverized coals is chosen specifically to illustrate the comparison of advanced magnetic separation systems such as might be developed for handling gas particulate separations in applications where the particle grain size is smaller than that presently considered practicable for physical coal beneficiation. These applications are significant and can include steel furnace off-gas cleanup as well as coal desulfurization. Extensions of the principles illustrated here may also find applications in electromagnetic control of entrained and fluidized bed reactors.

The fine-pulverized coal is pneumatically conveyed to the cyclone collector as shown in the upper left corner of Figure 3[5], where it is uncoupled from the gas and dropped through a lock hopper into a fluidized bed reactor where the coal is reacted with gaseous iron pentacarbonyl. The dry treated coal is passed through a second lock hopper at the bottom of the reactor and is pneumatically conveyed to a magnetic separator where clean coal and magnetic refuse are separated. Lock hoppering is required because of the toxic properties of iron pentacarbonyl and carbon monoxide.

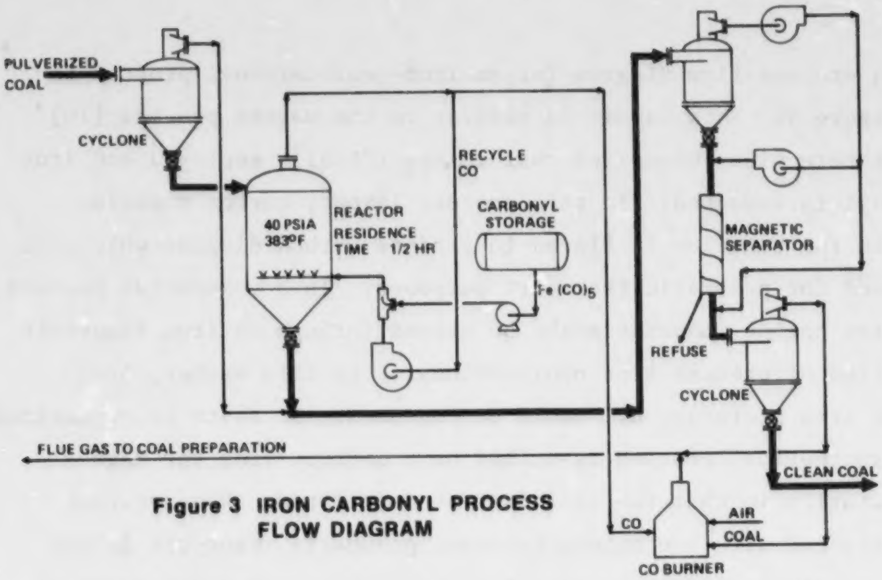

PULVERIZED COAL
CYCLONE
40 PSIA 383°F
REACTOR
RESIDENCE TIME 1/2 HR
RECYCLE CO
CARBONYL STORAGE
Fe (CO)5
MAGNETIC SEPARATOR
REFUSE
CYCLONE
CLEAN COAL
FLUE GAS TO COAL PREPARATION
CO
AIR
COAL
CO BURNER

Figure 3 IRON CARBONYL PROCESS FLOW DIAGRAM

Pulverized coal is pneumatically conveyed in this process with inert flue gas which is primarily carbon dioxide. Under normal operation conditions, between 2 and 5 lb CO_2 gas would be used to convey the coal at velocities betwen 2 000 and 5 000 ft/min. The use of these conditions would be prohibitive in the application of almost any type magnetic separator because: (a) very high velocity applications will require excessive pressure drop if a magnetized bed concept is employed, and (b) dilute solids suspensions will require large magnetized cross sections in order to be able to handle the full process flow. In order to circumvent these difficulties, a cyclone separator and bypass fans have been used to uncouple the coal dust from the main volume of entraining gas flow [8]. For the examples to follow, pulverized coal is conceptually blown through the magnetic separator at velocities ranging between the normal settling velocities for the dust and transport velocities of 5 000 to 6 000 ft/min. Gas loadings of four pounds of coal per pound of carbon dioxide are employed. The use of these conditions allows considerable latitude in the choice of magnet size.

Conceptual applications of batch-operated and smooth-bore magnetic separators have been compared for coals pretreated by the iron pentacarbonyl process described above. In the comparison, a high volatile A bituminous Pittsburgh Seam coal containing 1.2 wt.% iron pyrite was the coal (100 mesh top size and 50 micron mean particle size) that had been pulverized to 70 to 80% finer than 200 mesh (power plant grind). With this grind, mineral sulfur is assumed virtually completely liberated for this coal.

A. Batch-Operated HGMS-Type Separators

Magnetic capture of pneumatrically conveyed fine-pulverized "iron-rich disulfide" in HGMS-type separators has been calculated with the use of a mathematical model adopted from that developed to describe the capture of paramagnetic minerals from highly dispersed wet-minerals processes (kaolin clay) [14]. The model has been modified for this application to describe the capture of strongly magnetic materials from turbulent flows. The fractional removal of "liberated" magnetic particles is described by Equation (2) where C_0 is the ingress and C is the egress concentration of these materials.

$$R = \frac{C_0 - C}{C_0} = 1 - e^{-\frac{K_1}{CN_R/24}\left(\frac{1}{1+K_2}\right)} \qquad (2)$$

$$K_1 = \frac{4\pi M\rho_p \mu (d/D)^2 \varepsilon (1-\epsilon)L}{9\eta v} \quad \text{and,} \qquad (3)$$

N_R is the Reynolds number defined by,

$$N_R = \frac{\rho_f dv}{\eta} \qquad (4)$$

and C is the drag coefficient defined by the relationship,

$$\text{drag force} = 1/8 \rho_f C \pi d^2 v^2. \qquad (5)$$

At Reynolds numbers generally below 1/2, the drag force is given by Stokes Law,

$$\text{Stokes Law drag force} = 3\pi\eta dv. \qquad (6)$$

The term K_2 of Equation (2) is the ratio of particle weight to viscous drag,

$$K_2 = 4/3 \frac{\rho_p gd}{\rho_f v^2 C} \qquad (7)$$

and is generally unimportant for the flow conditions considered here.

The mathematical modeling of Equations (2)-(7) is for "ideal" unloaded filters and infers that retention in the magnetic filter is dominated by forces which hold the magnetic particle on the capture element. Capture on "loaded" filter elements can be handled empirically by a variety of methods including adjustments to the capture surface strand diameter D. In the above equations, M (gauss) is the magnetic capture surface magnetization, ρ_p is the magnetic particle density (gm/cc), μ is the "iron-rich disulfide" magnetic moment (emu/gm), d is the magnetic particle diameter (cm), D is the capture surface strand element diameter (cm), ε is the bed fractional void volume, L is the bed depth (cm), η is the fluid viscosity (poise), and v is the superficial velocity through the filter bed reckoned for 100% void volume, ρ_f is the fluid density (gm/cc), and g is the acceleration due to gravity (cm/sec^2). The fluidizing gas is carbon dioxide at 200°C where $\rho_f = 1.13 \times 10^{-3}$ gm/cc and $\eta = 2.27 \times 10^{-4}$ poise.

The magnetized bed configuration visualized for the conceptual application discussed here is a three-dimensional diamond-shaped structure composed basically of one-inch segments of 1 mm diameter No. 430 or No. 446 ferritic stainless steel. This material and its three-dimensional structure and element size have been chosen on the basis of commercial availability (demister tower packing), corrosive resistance under process conditions, immunity to abrasive wear, and effective magnetic removal of mean-size magnetic particles [8]. Coals pulverized to power plant fineness contain particles ranging downward from about 150 microns with a mean particle size of 50 microns (see Table II). A choice of thinner capture surface material, such as, for example, 100 micron diameter steel wool, would have provided for better removal of the finest size components of the coal dust, but its use would have involved more serious abrasive wear and would have led to increased pressure drop (in excess of 100 inches of water).

Because of the strong ferrimagnetism of the "iron-rich disulfide," a magnetic field of only 2 kOe has been employed. The magnetic field strength dependence of both the capture surface magnetization and of the calculated mineral magnetization are shown schematically in Figure 4.

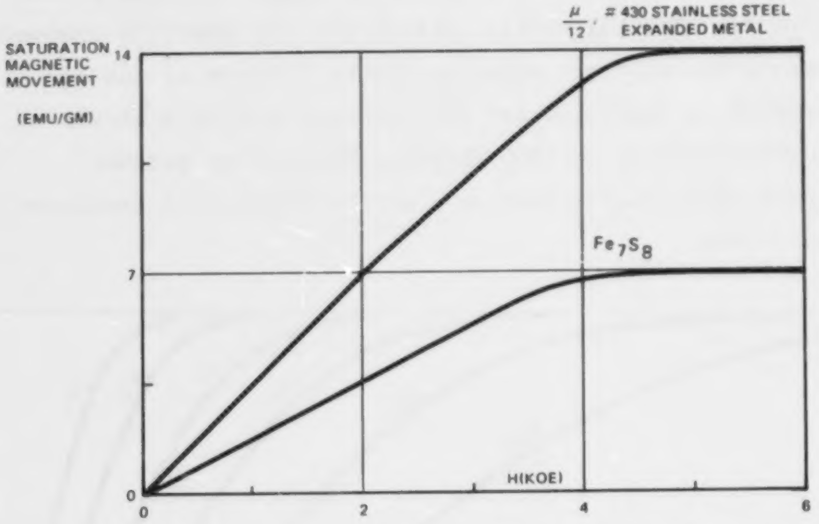


Figure 4 MAGNETIC PROPERTIES OF STAINLESS STEEL, AND Fe_7S_8

The ideal capture model of Equation (2) predicts virtually 100% removal of the coarse "iron-rich disulfide" particulates occurring in coals pulverized to power plant fineness. The broad performance characteristic inferred for HGMS removal of "iron-rich disulfide" from pulverized coal is shown in Table II.

TABLE II

Calculated HGMS Removal of 7 ∿8 emu/gm "Iron-Rich Disulfide" from Pulverized Coal

Particle Diameter (μ)	Particle Size Distribution (% Finer Than)	% Mineral Matter Removed Velocity (ft/min)	
		200	1600
150	100	100	100
100	83	100	100
50	50	100	100
10	18	100	100
3	7	100	99
1	3	99	45

On the basis of this simple model, it is theorized that HGMS-type filtration can achieve efficient removals of strongly magnetic mineral contaminants even at velocities which are large fractions of the entrained gas velocity. The calculated capture for fine particle components less than five micron diameter are shown in Figure 5, where it can be seen that 100% removal is predicted for particulates down to 3 micron diameter at flow velocities up to 1600 ft/min. Ninety-five percent removal of iron-rich particulates down to 4 micron diameter is predicted for flows of 6600 ft/min.

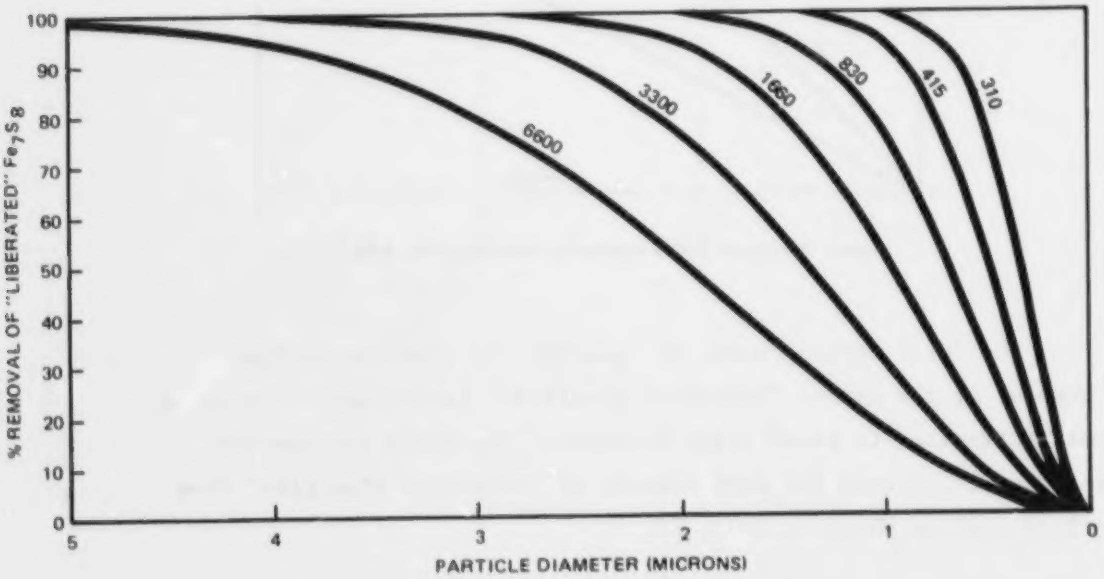


Figure 5 CALCULATED HGMS CAPTURE OF Fe_7S_8

The calculated capabilities discussed above suggest that HGMS has the potential for 90+% removal of mineral sulfur from fine-pulverized iron carbonyl treated coals in direct in-line application between the pulverizer and the burner. It is estimated that the HGMS-type separator can achieve this mineral sulfur removal for the reference coal with a 95+% recovery of mineral-free coal. This type process performance is not inconsistent with the experience of HGMS wet minerals beneficiation in the clay industry. Estimated add-on cost for desulfurization alone with the iron-pentacarbonyl process described here (including batch-operated HGMS separators) is less than $7.00 per ton of clean coal and compares favorably with current costs for high level commercial physical coal cleaning [15]. Conventional utility coal washing, however, is generally limited to the processing of coals coarser than 35 mesh.

1. Magnet Structure

The pillbox-type HGMS magnetic structure shown in Figure 1 is not well suited to high velocity processing. The top and bottom of the HGMS structure of Figure 1 form the north and south poles of the magnet and are solid iron pole pieces. In the wet application, slurries enter and exit the magnetized bed through a multiplicity of small diameter holes bored through the pole pieces. The entrance and exit pipes also promote flow distribution in the relatively dense mineral slurry. Complete utilization of the entire magnetized volume is essential to the economic viability of the HGMS concept, so that good flow distribution and low pressure drop are essential features of the design of entrance and exit to the magnetized bed.

It is believed that neither good flow distribution nor low pressure drop can be achieved with the magnet structure shown in Figure 1 in high velocity gas-particulate separations. Use of an iron-clad Bitter solenoid [16] with poles removed or of a cross-flow H-frame structure recently proposed by Allen [17] would provide better means for flow distribution and pressure drop control than does the pillbox magnet structure of Figure 1. The Bitter-type magnet, with iron poles removed, would allow for ease of access to the magnetized volume but would have marked disadvantages of requiring higher currents to achieve fields equal to those developed by the magnet structure of Figure 1 and would have the added occupational hazard of large dipole fringing fields outside the magnet in the working environment.

The magnet conceptualized for the gas-particulate separator is designed around the H-frame magnet structure proposed by Allen and shown in Figure 6. This magnet is essentially identical to the pillbox magnet of Figure 1, except that the iron on two of the outside edges has been removed, and the energizing coils have been separated in the center and folded up in a saddle shape so as to permit transverse access to the magnetized volume. Performance characteristics for this type magnet are essentially identical to those of the Bitter-type magnet structure, but dipolar fringing fields are low.

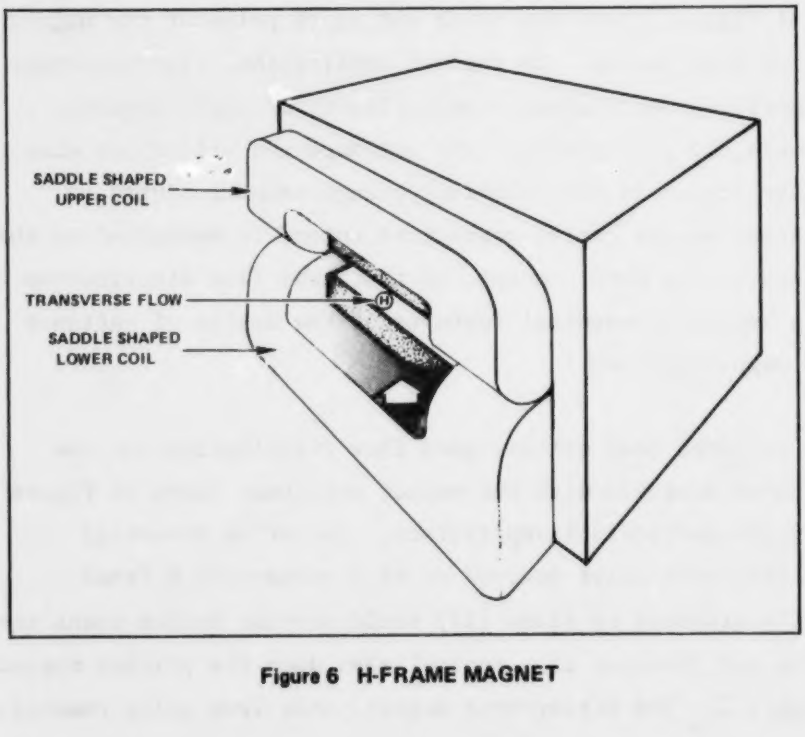


Figure 6 H-FRAME MAGNET

The H-frame magnetic separator is different from the pillbox structure in another important aspect. In the pillbox structure, both filtration flow and flushing flow occur along the same length which is parallel to the magnetic field direction in Figure 1. Flushing can be difficult when the pillbox separator employs a deep filter bed. The Allen patent teaches filtration flow along the long pole length transverse to the short pole gap opening and flushing flow, with the magnet de-energized, along the short length parallel to the pole opening. In this fashion, flushing time is expected to be less, for equal bed volume, than for the pillbox separator. Reduction of dead time in filter cleaning is important in improving the batch-operated filter on-line time.

With the use of transverse flow the relative orientation of magnetized capture surface element, magnetic field direction, and flow velocity is different from that employed with the pillbox magnet structure. It is believed, however, that the magnetic capture surface designed for this study permits good capture of strongly magnetic materials at low pressure drop in high velocity turbulent flow applications.

2. Process Characteristics

Magnet size to handle process capacity is determined with use of Eq. (8).

$$C = 0.72\ S_c E v A \varepsilon \qquad (8)$$

In Eq. (8), C is the process capacity (8 000 tons coal per day), S_c is the pounds of entrained coal per cubic foot of flue gas (0.28), E is the on-line factor for the batch-operated magnet, v is the flow velocity in the filter bed (ft/min), A is the cross-sectional area of the filter bed (ft^2) for a magnet operating with on-line factor E, and ε is the filter fractional void volume.

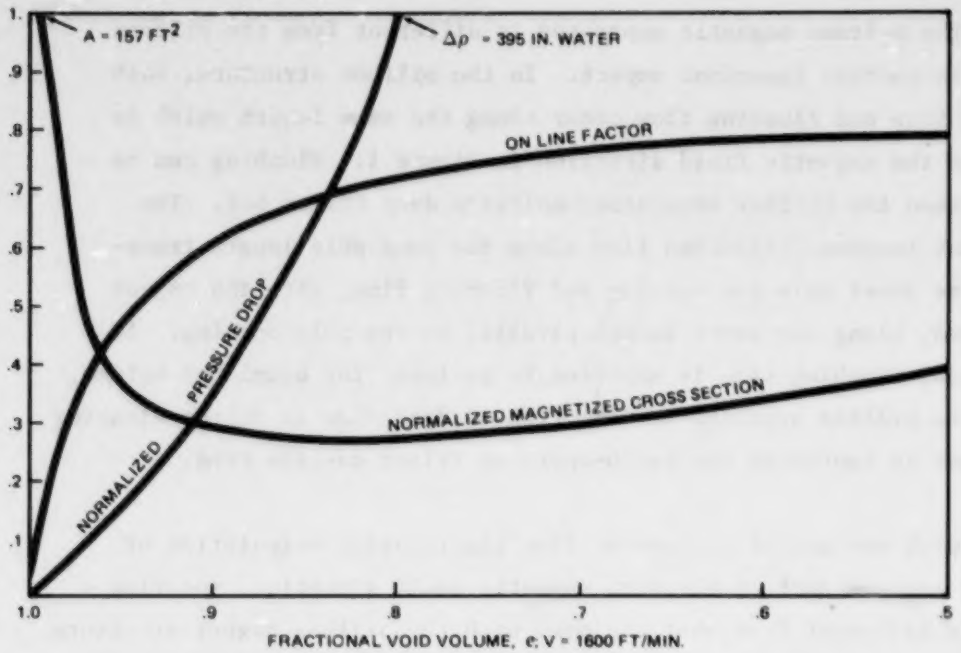


Figure 7 EFFECT OF BED PACKING DENSITY ON MAGNET SIZE AND PRESSURE DROP AT 1600 FT/MIN. GAS FLOW

Calculations of magnet size requirements (magnetized cross-sectional area) and pressure drop across the filter bed are shown in Figures 7 and 8 as functions of the bed fractional void volume and velocity of flow through the bed, respectively. The calculation of magnet size involves use of the on-line factor shown in Figures 7 and 8. Details of on-line factor calculations can be found in the literature [18]. As is apparent in the figures, significant reductions in magnet size requirements can be achieved with use of densely packed filter beds, but only at the expense of high pressure drop. By way of example, a single 2 kOe HGMS-type magnetic separator operating at 60% on-line factor could handle the entire 8 000-ton-per-day process stream with an 8.3-ft deep bed of 45 ft^2 cross-sectional area when packed to 90% void volume. The device would develop a calculated pressure drop of 138-inch water gauge. At a flow rate of 39 000 cfm, the system would require a 1400 hp auxiliary fan of 1.05 megawatts peak power consumption. This power requirement is large compared with the 140 kilowatts estimated for magnet operation.

High velocity gas entrained flow of particulates through porous media is a very complex phenomenon, and magnetic capture in such beds will probably have to be studied and understood on an individual basis. The calculations presented illustrate the important point that the magnetic separator cannot be isolated from the overall process for planning and estimating purposes because choice of auxiliary process operations and conditions may well be more important than magnet optimization.

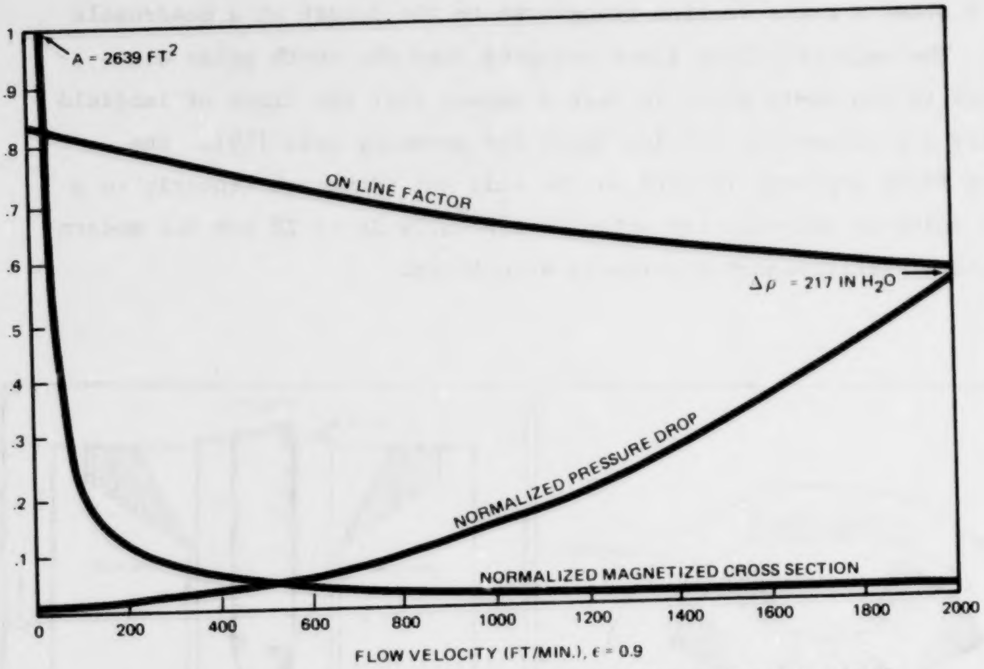
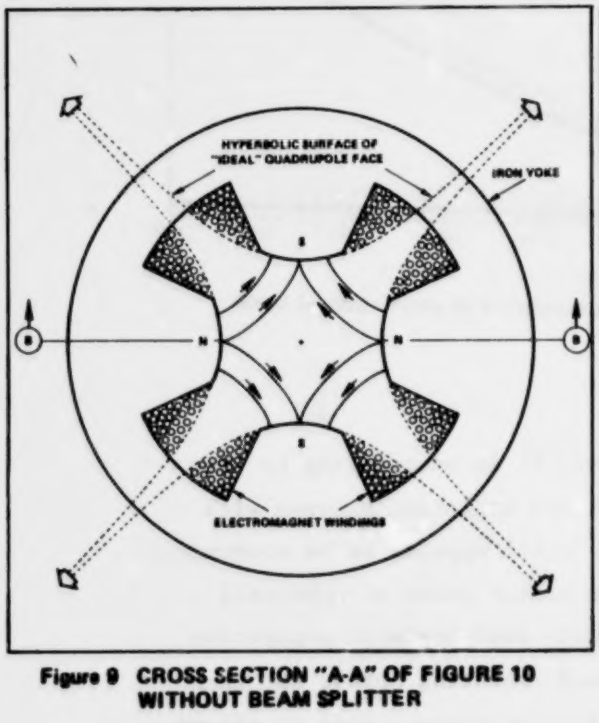


Figure 8 EFFECT OF FLOW VELOCITY ON MAGNET SIZE AND PRESSURE DROP AT 90% VOID VOLUME

While the example presented is a severe one, it is encouraging to know that the use of HGMS hypothesized for this dry application, even with its expensive ancillary cyclones and fans, still appears to be economical in the coal processing context. Estimates of the costs of HGMS coal cleaning with the Magnex process [5] indicate that magnetic separation will be a small component of the capital and operating costs, and that this type of coal cleaning can be expected to be accomplished at add-on fuel costs for desulfurization of about $7.00 per ton of clean coal produced. This estimated cost is competitive with that known for stack gas desulfurization and with those estimated for more elaborate chemical cleaning and coal conversion technologies.

B. Quadrupole Magnetic Separators

Quadrupole magnets are structures which produce a uniform magnetic field gradient throughout a relatively large working volume. Figure 9 shows a cross-section transverse to the length of a quadrupole magnet. The magnetic field lines emerging from the north poles are deflected to the south poles in such a manner that the lines of isofield intensity are concentric circles about the symmetry axis [19]. The magnetic field strength is zero on the axis and increases linearly to a maximum value on the pole tip which is typically 10 to 12 kOe for modern iron-return electromagnet quatrupole structures.


HYPERBOLIC SURFACE OF
"IDEAL" QUADRUPOLE FACE
IRON YOKE
S
N
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ELECTROMAGNET WINDINGS

Figure 9 CROSS SECTION "A-A" OF FIGURE 10 WITHOUT BEAM SPLITTER

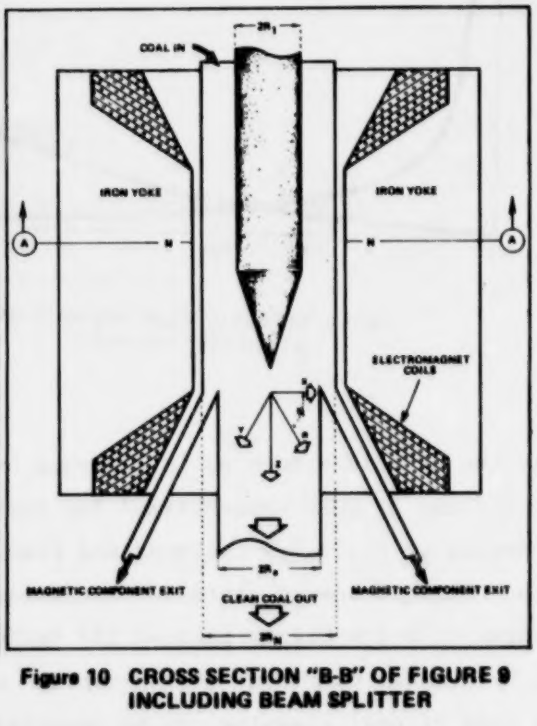
2R1
COAL IN
IRON YOKE
IRON YOKE
A
N
N
A
ELECTROMAGNET COILS
X
N
Y
R
Z
2R0
MAGNETIC COMPONENT EXIT
CLEAN COAL OUT
MAGNETIC COMPONENT EXIT
2RN

Figure 10 CROSS SECTION "B-B" OF FIGURE 9 INCLUDING BEAM SPLITTER

Figure 10 shows a vertical section through the symmetry axis of a quadrupole structure in which a "beam-splitter" pipe has been inserted for the purpose of separating the clean coal product from the magnetic fraction. The coal enters the quadrupole field region in the annular space of inner and outer radii R_I and R_N, respectively. The inside radius R_I is the minimum distance from the symmetry axis for which the magnetic field strength is sufficient to saturate the magnetization of the "iron-rich disulfide."

$$H(R_I) = H_S = \frac{R_I}{R_N} H_O \qquad (9)$$

The outer radius, R_N, is chosen so that N magnets with annular entrance areas, $\pi(R_N^2 - R_I^2)$, can handle the entire process flow

$$C(\mathrm{lb/min}) = \frac{\rho_{CO_2} N\pi(R_N^2 - R_I^2)V_o}{W} \qquad (10)$$

where W is the carbon dioxide to coal ratio, ρ_{CO_2} is the CO_2 gas density ($\mathrm{lbs/ft^3}$), and V_o is the gas flow velocity (ft/min).

In flowing through the quadrupole field region, the magnetic component will be deflected radially outward toward the magnetic poles and will exit through a take-off pipe on the outside diameter of the entrance pipe. A small fraction of the clean coal inadvertently exists with the magnetic component, but the major portion of the clean coal emerges from the pipe of radius R_e at the bottom of the quadrupole beam splitter.

The equation of motion along the vertical axis is

$$\frac{dZ}{dt} = V_O + \frac{2\rho r^2 g}{9\eta} \qquad (11)$$

where ρ is the particle density, r is the particle radius, g is the acceleration due to gravity, and η is the fluid viscosity. Buoyancy forces have been neglected because of the great difference between particle and gas densities. Stokes Law has been assumed for the viscous drag force, and particle-particle interactions have been ignored. The residence time, T, in the magnet structure of length L is given by the relationship

$$T = \frac{L}{V_o + \frac{2\rho r^2 g}{9\eta}} \tag{12}$$

With similar assumptions, the radial component of the equation of motion is given in Eq. (13), and the time-dependent radial motion is given in Eq. (14) where the particle is assumed to have entered the field region on initial radius

$$R_I \leq R_i \leq R_N$$

with zero radial velocity.

$$m_p \ddot{R} = \mu m_p \frac{\partial H}{\partial R} - 6\pi\eta r\dot{R} \tag{13}$$

$$R(t) = R_i + \frac{\gamma}{\alpha^2} (e^{-\alpha t} - 1) + \frac{\gamma}{\alpha} t \tag{14}$$

In Eq. (14), $\gamma = \frac{\mu H_O}{R_N}$ where μ is the "iron-rich disulfide" ferrimagnetic moment,

$$\alpha = \frac{6\pi\eta r}{m_p} = \frac{9\eta}{2\rho r^2} \tag{15}$$

and H(R) is equal to $H_O R/R_N$.

Assuming that all particles which strike the inner wall at radius R_N are captured, the equation of motion, Eq. (14), and the residence time, T, can be combined to implicitly define a quadrupole-structure length, $L_N(r,p,V_O)$, such that all particles of radius greater than r will be collected where p is the percent of such particles in the annular region between R_i and R_N.

$$L_N(r,p,V_O) = \frac{KT}{NR_N^2}\left(1 + \frac{gNR_N^2}{K\alpha}\right) \tag{16}$$

where

$$K = \frac{W}{\pi\rho_{CO_2}[1 - (H_S/H_O)^2]} \tag{17}$$

and

$$C = \frac{\pi\rho_{CO_2}\, NR_N^2 V_O}{W}\,[1-(H_2/H_O)^2] \tag{18}$$

In these equations, T is the solution to the equation f(T) = 0 where

$$f(T) = e^{-\alpha T} + \alpha T - \alpha\left[\frac{\alpha R_N^2}{\mu H_O}\,\{1-\sqrt{1-[1-(H_S/H_o)^2]P/100}\} + 1/\alpha\right] \tag{19}$$

The appropriate parameters are: $\rho_{CO_2} = 1.129 \times 10^{-3}$ gm/cc, $\eta_{CO_2} = 229 \times 10^{-6}$ poise, $\mu = 6.9$ emu/gm, $H_O = 10\ 000$ Oe, $W \simeq .25$ lb CO_2/lb coal, and $\rho \sim 4.5$ gm/cc for "iron-rich disulfide."

Figure 11 shows the calculated length requirements for a quadrupole magnetic separator which achieves 100% removal of 10 micron diameter Fe_7S_8 from pneumatically conveyed pulverized coal. Performance at this level is equivalent to 90% removal of all mineral sulfur from the coal.

It can be seen that a quadrupole structure is not as efficient in removal of fine-sized particles as is HGMS. This is basically because of the low magnetic field gradient, 656 gauss/cm, produced by the quadrupole. Unfortunately, the quadrupole magnet requires full field strength operation to produce this field gradient even when processing strongly magnetic materials. With use of the above idealized equations, it can be seen that a quadrupole of about 16.5 ft length by 1 ft diameter opening will be required for 100% removal of 20-micron diameter magnetic particulates from a 8000 ton coal per day process stream flowing at velocities of 1600 ft/min.

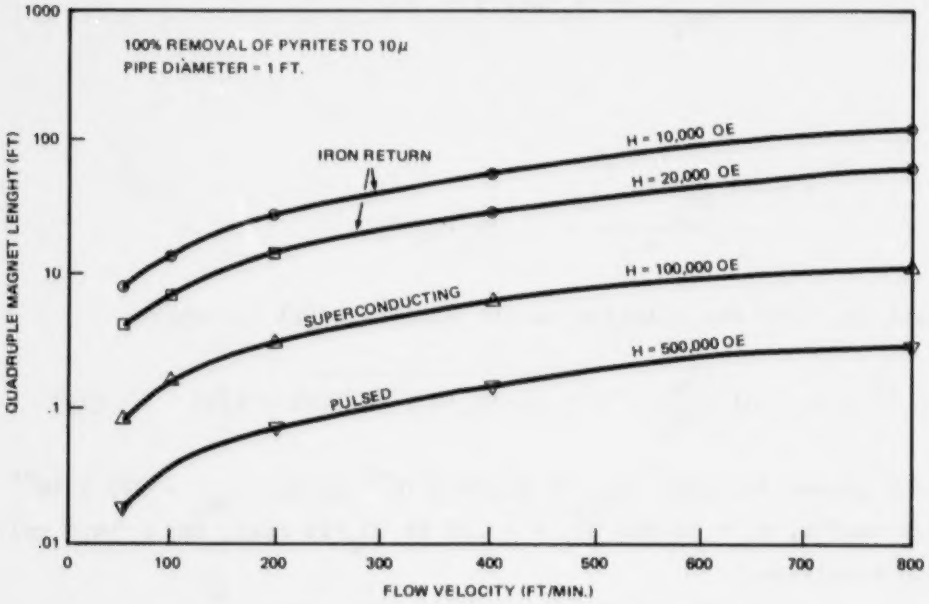

Figure 11 QUADRUPLE LENGTH REQUIREMENTS

It is interesting to note the significant reduction in quadrupole lengths that could be achieved if increased field strengths could be utilized. With pole tip fields of the order of 100 kOe, quadrupole magnetic separators would begin to be competitive with HGMS structures in this application.

1. Magnetic Structures

Quadrupole magnet costs and power consumptions have been calculated from design bases developed at the Lawrence Berkeley Laboratory [20]. It is of interest to note that there is a power/field-gradient trade-off implicit in the design of quadrupole magnet structures.

In Figure 9, the surface shape of the "ideal" pole for the quadrupole structure is shown by the dotted lines. To make the "real" quadrupole, shown by the solid line, the iron of the "ideal" pole pieces has been shaved back in order to make room for the electromagnet windings. The windings are shown as circles in the figure. Because of the relatively small space made available in this shaving operation, the conductor cross-section per winding is limited so that quadrupoles are inherently power-consuming devices.

A conservative design basis has been chosen whereby moderate pole piece field strengths can be developed without excessive power requirements (Modification III, pages 8 and 9, Reference [20]).

2. Process Characteristics

It has been noted that smooth bore magnetic separators will be inherently low pressure drop devices. By way of comparison, it can be calculated that the power requirement of all fans to supply the quadrupole magnet installation to handle 8000 T/D coal is only about 100 hp as compared with 1400 hp for HGMS desulfurization. While quadrupole particulate removal performance may not equal that of the batch-operated device, it well could prove to be the case that ancillary process requirements might dictate cost/benefit-type compromise in determining overall process optimization.

CONCLUSIONS

A technical and cost comparison for HGMS and quadrupole magnetic separators is shown in Table III. The HGMS installation is estimated to be about a factor of 5 less expensive than the bank of 33 quatrupoles and consumes about one-seventh the power. If auxiliary fan requirements are now included, however, the HGMS installation cost estimate increases to about one-fourth that of the quadrupoles, and the total power consumption estimate increases to about one-half that of the quadrupole installation. Including fan requirements, operating costs estimated for the HGMS installation are only about one-half that of the conventional iron-return-electromagnet quadrupole structure (power assumed at 33 mills/kW-hr).

Table III

Conceptual Level Comparison of HGMS and Quadrupole Magnetic Separators for Dry Coal Desulfurization

	HGMS	Quadrupole
100% Removal Size (μ-esd)	3	20
% Removal - All Pyrites	96	82
% Coal Recovery	98	97
V (fpm)	1600	1 600
No. Magnets	4	33
Duty Cycle (%)	50	100
Field Strength (Oe)	2000	10 000
Total Cost ($)		
All Magnets	327 000	3 123 000
Canisters	175 000	19 000
Matrices	97 000	0
Power Supply (including 15% uncertainty)	28 000	201 000
	627 000	3 343 000
$/ton Clean Coal	0.03	0.13
Installed Capital Cost / Actual Process CFM	$1.22/CFM	$6.6/CFM

It is now interesting to again recall the results of Figure 11 which suggest that if superconducting or pulsed field quadrupole magnets could be developed to significantly increase pole tip field strength, then quadrupole units would be more competitive with HGMS on a cost-effective basis. While power consumption could be expected to be significantly lower for a superconducting magnet installation, for example, capital costs cannot be assessed at this time because there is no history of commercial superconducting magnet installations. A factor of 10 increase in field strength will result roughly in a factor of 10 reduction in pole length, but the economy of scale in superconducting installations is now known. If capital costs were reduced by a factor of four, operating costs for the two approaches would be similar. The point is that one cannot properly estimate magnetic separator costs alone for comparative purposes but must estimate the separator costs within the process context.

The mathematical model of capture in the HGMS device that has been employed in this study is a straightforward modification of a model successfully developed for describing the capture of weakly paramagnetic materials in the wet beneficiation of kaolin clay. The modelling reported here has, accordingly, assumed a modest surface loading of 30 wt%. While this assumption may be appropriate to applications such as coal desulfurization, where nonmagnetic particulates, in particular fine clay components, are inadvertently mechanically captured, it may be overly conservative in applications such as steel furnace off-gas scrubbing where particulate concentration is very low, but where the percentage of magnetic particulates is high.

In the magnetic capture of very fine size and strongly magnetic materials, the continued capture of fine particulates may even be enhanced by the history of fine particulate capture--up to a point. There are local surface imperfections on the capture material which produce highly localized capture forces distributed over the entire surface of the magnetized bed which are believed to be selective to micron and submicron size particulates. Very fine size and strongly magnetic particulates which are captured at these force centers may have a tendency to create new sites for further fine particulate capture. Turbulent flow aids this process in bringing the particulates to the capture surface from all directions.

Paramagnetic particles, on the contrary, are believed to be captured preferentially on the leading and trailing edges of the magnetic capture material in the regions of stagnation points for laminar flow [14]. Magnetic filter performance degrades continuously with time in the processing of these materials (kaolin clay) because of the extremely weak forces for capture of paramagnetic materials. This loss of efficiency is due to particle buildup on the capture surface. In effect, the capture of paramagnetic material destroys surface area for collection. By way of contrast, the capture of very fine size but strongly magnetic materials creates rather than destroys new capture sites. This being the case, one would expect HGMS filter performance to be relatively uninfluenced by time in the capture of strongly magnetic materials until such time as a breakthrough occurs due to turbulent breakup of the captured coagules.

The magnetic desulfurization of coal described here illustrates an extreme and difficult application of magnetic separation technology. In this example, particle size distribution is very broad, the flow velocity is very high, and restrictive pressure drip requirements are imposed by process considerations. The performance characteristics calculated for both HGMS and quadrupole separation indicate an attractive possibility for the use of magnetic methods in gas-particulate separations for strongly magnetic particulates. Some very interesting magnetic separator development lies ahead.

ACKNOWLEDGMENTS

The results reported here are a continuation of work originally supported by the U.S. Bureau of Mines. It is a pleasure to acknowledge technical assistance provided by Messrs. F. A. Karlson, J. Vicory, and John McCaig, and to thank Messrs. G. H. Dyer and E. L. Ekholm for making Bechtel resources available for this investigation.

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WRITTEN DISCUSSION

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Dr. Oder has presented an interesting scheme for processing dry coal by HGMS. I would like to stress however, that there is really no experimental data at this time to support the implicit assumption that the separation of pyrite from coal suspensions in high velocity air streams will be as straightforward as the separation from water slurries. We have done a little work on this problem and can report that the problem is more complex. Coal which can be successfully processed in water slurries by HGMS does not respond when simply propelled through a matrix in high air stream. Agglomeration and stickness appear to be major problems. It may be that pretreatment of the coal to enhance the magnetic susceptibility of the pyrite may help overcome this difficulty but this is yet to be established.

Successful dry magnetic separation of coal was reported ten years ago by Kester et al.;[1] they were most successful with an Upper Freeport coal, removing 80% of the pyritic sulfur. They used a Frantz Isodynamic Separator which is not a high gradient device but rather a laboratory instrument for making a split between a magnetic and a non-magnetic fraction. The mineral particles slide down a tilted inclined plane. The magnetic force competes with the gravitational force to separate the particles into separate streams. The particles move rather slowly under the influence of these two forces and are not air propelled. It is essentially an open gradient device and therefore in the same general category as the quadrupole separator but substitutes gravity feed for air propulsion. At the Auburn Conference on coal desulfurization by HGSM, H.H. Murray[2] reported on some HGMS separation of dry coal again using gravity feed.

Whatever the other problems of magnetically desulfurizing coal may be, the conversion of pyrite to pyrhhotite is a forward step if it can be carried out economically. It would be useful to have more data on the reaction kinetics of the conversion process. The magnetization of the FeS_x sulfides is a sensitive function of the composition parameter x, as illustrated in the figure. In his calculation of the magnetization of the converted pyrite Dr. Oder assumes the end product is simply Fe_7S_8. Actually a continuous range of compositions is possible; the other stoichiometries have lower magnetization. Incidentally, the composition $Fe_{1.44}S_2$ does not correspond to Fe_7S_8.

The temperature dependence of the magnetization is quite complex for some of the sulfides, as shown in the work of Lotgering.[3] From his investigations it appears that rapid quenching of the pyrhhotite from 200°C. to ambient temperature should result in a significant increase in the magnetization. The separation should therefore be carried out in a cooled air stream and not at the Reaction temperature.

Dr. Oder mentions that because of the toxic properties of iron pentacarbonyl a lock hopper is required to contain it. How good must the lock hopper be? What is the experience in handling the relatively large quantities of similar toxic materials under conditions simulating power plant operation? Power plant operators are generally reluctant to undertake operations which require chemical engineering expertise because theirs is a different kind of operation.

Dr. Oder points out that the power consumption of the fans required to deliver the air for propelling the coal is a significant factor in the overall cost. Is it worth considering alternate schemes for carrying out the beneficiation? For example, one might visualize a process in which the separation was carried out in a separator by gravity feed and the coal was then pelletized. In that case the processing could be done near the mine and the power plant would be saved the task of disposing of the reject material and paying for its freight.

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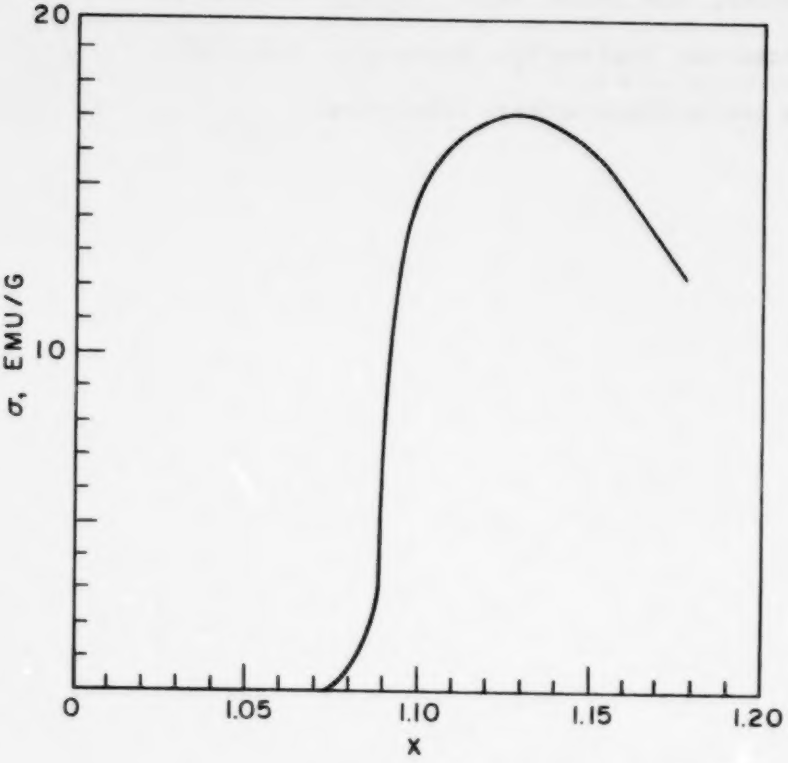
*Supported by the National Science Foundation.


20
10
0
σ, EMU/G
1.05
1.10
1.15
1.20
X

OPEN DISCUSSION

Oder: Dr. Maxwell mentions, for reasons other than what I had pointed out, that the straight use of the magnetic separator in the processing will not occur. That is true. I don't know whether it has to do with agglomeration or not. Our point simply is that, just from an economic standpoint, you need a 20 kilo horsepower magnet to handle the process strain. Those are expensive to make. From that standpoint alone we do not think it is economical, and suggest looking at alternative approaches such as Hazen . There are chemical and thermal methods that do the conversions well. With regard to the agglomeration problem it is my hunch that turbulent flow through that device actually aids magnetic separation. I can not prove it, but it aids it in two regards. First is the stripping that was mentioned. This is a very difficult gas - particle separation. Because you have got a surface for capture and therefore you can capture magnetic particles. You can also mechanically entrain non-magnetic particles. This is to your disadvantage with HGMS technology because you eat up captured surface. If you have got a turbulent drag force that carries those non-magnetic particles out of there, then you don't hinder your captured surface characteristic which is good.

The other point is that when you capture a strongly magnetic particle on a captured surface, if that particle is itself equivalent in magnetization to the capture surface, then you create more capture surface. The turbulent flow will bring particles in to the new sight from all directions. Therefore the probability of capture is high.

Again the feeling is that processes like Hazen are good. They have done this at Debang. They have a big scale unit that is running. They also have a pilot unit that has been running for four months. In that four month period, they have brought non-compliance coal into compliance and they have done this basically so far with coarsely pulvarized coal (only going to 14 mesh). They select their coal so the pyrites have a coarse size distribution. They have been clever in that regard. They choose Upper Freeport coals or particular coals that are pyrite, they respond well, and don't have to be pulverized. My point is that, if you are going to look at mid-western and eastern coals, you are going to have to face the problem of pulverizing it. To do that you are going to have the technology to handle pulverized coals. I don't think it will work unless you convert to pyrites.

Maxwell: On the role of turbulence, I think if you manage to capture the pyrhhotite, since it is more magnetic,you have a chance to strip any coal attached to it by turbulence.

I don't think captured pyrhhotite particles will act very effectively as nuclation centers simply because the magnetization of pyrhhotite is not that large. We have seen, for example, studies of captured magnetite and hematite particles. Magnatite particles have a magnetization saturation of about one-third that of iron and act as a nucleation center. Drum separators are an example. Hematite on the other hand, which is probably down in order of magnitude from magnetite, is perhaps probably more comparable to pyrhhotite and doesn't behave in this way. It collects until it shields the captured particles from the rest of the other particles then you no longer capture particles. So I don't really think that pyrhhotite particle would increase your effective capture.

Friedlander: At higher velocities we have more problems, strangely enough, with liquid systems and we get erratic behavior. You can make matters worse by increasing velocity to the region of unsteady flow.

Baril: You are proposing to do a process that people demonstrated could be done electrostatically many years ago. That process has not been successful or applied. The reason is that not all coals are the same. The objective is to remove the sulfur not just the pyrite. Sulfur occurs both in the form of pyrite and organically formed sulfur which is not released by grinding Incidently all pyritic sulfur is not released by grinding in some coals. Generally you grind out to 200 mesh to relieve most of it in most coals. If you don't remove all the sulfur, you have to process most coals through some other operation to meet the emission standards. So actually you have not accomplished anything more than has been accomplished 10 or 15 years ago.

Oder: That is not true. We have studied the chemical processes in great detail. We studied the TRW process and the whole schlemiel. The idea is to look at the limit on the organic sulfur level that you think you must remove to get a significant upgrading of our Eastern reserves. You take a look at that and the cost of chemical cleaning, projected at $20 per ton, this implies that you must come back to physical cleaning and do a better job of it, because you are talking about dollars per ton. You have to draw a line in there somewhere. There are some coals that can be cleaned well physically, some chemically and some converted through gasification and liquefaction. But there is no single technology that is going to handle it all.

Baril: Electrostatic techniques require less power than magnetic techniques. There are existing large scale separators in operation like the Wabush in Canada. They are removing magnetite from sand and upgrading iron work; it is commercial. What is the advantage of going to magnetic if there is another process which is perhaps cheaper? It may be more expensive for initial installation costs; I do not know. But the operation costs would be lower.

Liu: On behalf of all the magnetic people, I would like to defend your criticism about comparison of magnetics and other approaches. There is various documentation I can show you that your argument is not necessarily true. Over 17% of the annual coal production in the U.S. is low enough in organic sulfur and less than 0.7% of that can't be cleaned magnetically. It costs less than all the chemical or physical processes.

MAGNETIC SEPARATION OF PARTICULATE AIR POLLUTANTS

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INTRODUCTION

In the last several years, particulate control technology has been developed to the extent that methods are now available to control large particle emissions from most industrial sources with an efficiency of greater than 99 percent. Emphasis on particulate control has now shifted toward the fine particle size range, particularly to particles which have diameters between 0.3 and 3 micrometers. These particles are of primary interest because they tend to remain in the atmosphere for long periods of time where they may cause atmospheric haze and be transported to cause a long range adverse health impact. They also are the particles which readily enter and deposit in the human respiratory system.

The three major conventional technologies for the control of fine particulate emissions are electrostatic precipitation, wet scrubbing, and fabric filtration. The cost of applying these control methods and the fractional collection efficiency obtained vary according to the specific application. Recent EPA research in particulate emission control has followed two complementary courses: (a) study of the fine particle collection characteristics of the conventional technologies and (b) evaluation of viable alternatives to the conventional technologies. Both of these courses have as their goals the enhancement of fine particulate collection efficiency and the reduction of the capital and operating costs of industrial emission control.

In the last decade research and commercial applications have demonstrated that high gradient magnetic separation (HGMS) is an effective and economical method of removing small paramagnetic particles from selected liquid streams. In this study, published theoretical and experimental reports of HGMS were utilized to evaluate the potential success of the process in removing fine, paramagnetic particles from gas streams. Industrial sources of fine, paramagnetic particles were identified, and the steelmaking basic oxygen furnace (BOF) was selected as the primary candidate for initial HGMS evaluation. A bench-scale apparatus was constructed and HGMS experiments were conducted using redispersed BOF dust. With existing theoretical equations as a guideline, the experimental data were analyzed and favorable operating conditions were identified.

THEORETICAL ASPECTS OF HGMS

To evaluate the usefulness and limitations of high gradient magnetic separation, the theoretical basis of the concept must be examined. Engineers have long sought a simple but accurate theoretical model to describe conventional filtration. The mathematics of most useful models quickly becomes unwieldy, necessitating simplifying assumptions. Nevertheless, the solutions often yield valuable information about the phenomenological effects of key operating variables.

The recent development of HGMS models has followed much the same course. Watson has published two papers [1,2] describing the single-fiber model depicted in Figure 1. The radius of the cylindrical ferromagnetic wire is a, and the radius of the particle is R. For r>>a, the velocity of the particle is V_0 and parallel to but in the negative x direction.

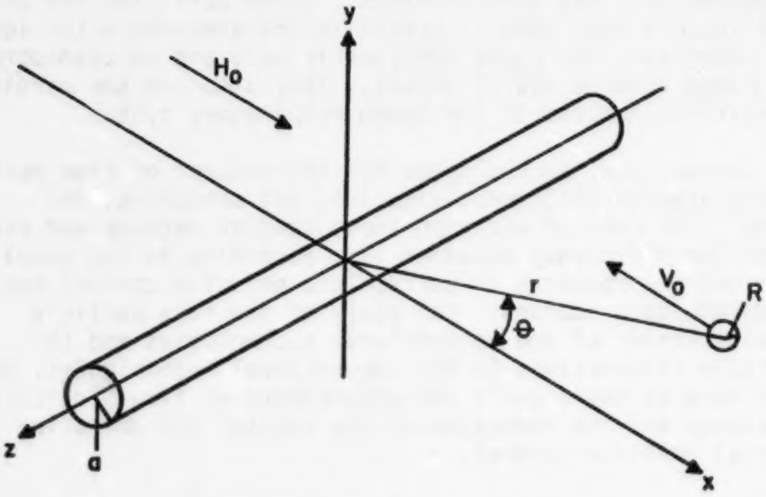

Figure 1. Coordinate system for HGMS single-fiber model.

For r>>a, the magnetic field strength is H_O and parallel to but in the positive x direction. Near the wire the field is concentrated, and a net magnetic force acts on the particle, changing its trajectory. Watson developed the expression for the magnetic force and incorporated it in the equations of motion to describe the particle trajectory.

Watson presented an argument that the inertial force terms in the resulting equations are negligible in comparison to magnetic and viscous force terms [1], and finally obtained for the radial component of motion

$$\frac{dr_a}{dt} = \frac{V_o}{a}\left(1 - \frac{1}{r_a^2}\right)\mathrm{Cos}\Theta - \frac{V_m}{a}\frac{1}{r_a^3}\left(\frac{M_s}{2H_o r_a^2} + \mathrm{Cos}2\Theta\right) \qquad (1)$$

and for the angular component

$$r_a \frac{d\Theta}{dt} = \frac{V_o}{a}\left(1 + \frac{1}{r_a^2}\right)\mathrm{Sin}\Theta - \frac{V_m}{a}\frac{\mathrm{Sin}2\Theta}{r_a^3} \qquad (2)$$

where $r_a = \frac{r}{a}$, normalized radial position, dimensionless

a = radius of wire, m

t = time, s

V_o = fluid velocity, m/s

Θ = angular position

$V_m = \frac{2\,\chi\, M_s B_o R^2}{9\,\eta\, a}$ magnetic velocity, m/s

where χ = magnetic susceptibility of particle, dimensionless

M_s = saturation magnetization of wire, A/m

B_o = magnetic flux density, T

R = particle radius, m

η = fluid viscosity, Ns/m^2

H_o = magnetic field strength, A/m.

The fluid velocity is assumed to be normal to the wire at some point upstream where the x-coordinate is large compared to the wire diameter. The solution of equations (1) and (2) can then be used to identify particle trajectories which result in the particle striking the wire as illustrated in Figure 2. The coordinates are normalized by the wire radius: the initial coordinate corresponding to the maximum y whose trajectory results in collection, $(\frac{y}{a})_{max}$, is defined as R_c. All particles with initial coordinates $(\frac{y}{a})_i$ such that $-R_c < (\frac{y}{a})_i < R_c$ will be captured. The dimensional capture cross section of the wire is thus $2R_ca$. For $\frac{V_m}{V_o}$ less than 10, R_c is also a function of the dimensionless quantity $\frac{\mu_o M_s}{2B_o}$ where μ_o is the permeability of free space. This paper references two situations: $\frac{\mu_o M_s}{2B_o} < 0.2$ (Case I) and $\frac{\mu_o M_s}{2B_o} \simeq 2$ (Case II).

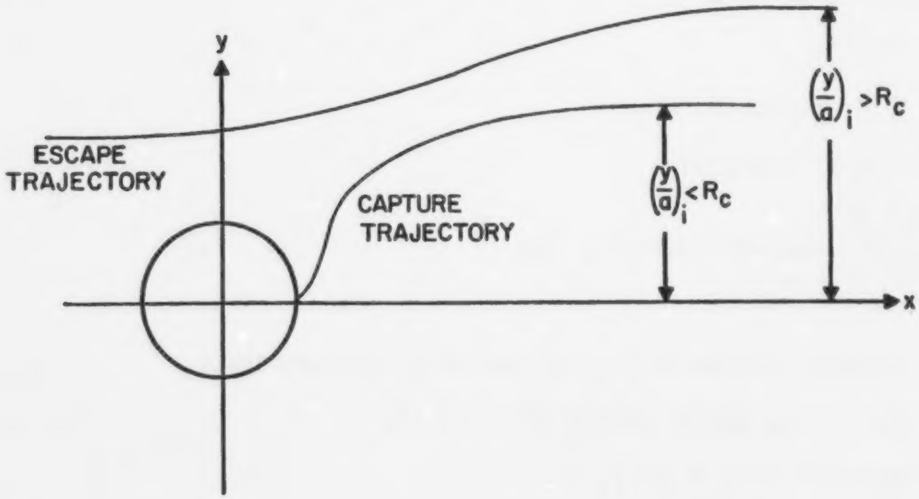
y
ESCAPE TRAJECTORY
CAPTURE TRAJECTORY
$\left(\frac{y}{a}\right)_i < R_c$
$\left(\frac{y}{a}\right)_i > R_c$
x

Figure 2. Illustration of particle trajectories under the influence of HGMS.

To calculate the collection efficiency for an HGMS apparatus containing many randomly oriented cylindrical fibers, Watson [1] first extrapolated from the single fiber to a filter element of differential thickness and unit cross-sectional area as follows:

$FAdL$ = total volume of wire,

FdL = volume of wire/unit cross section of filter enclosure,

$\frac{FdL}{\pi a^2}$ = length of wire/unit cross section of filter enclosure,

$\frac{FdL}{\pi a^2}(2R_c a)$ = total capture cross section/unit cross section of filter enclosure.

where F = volumetric packing density of mesh, dimensionless

A = cross-sectional area of the filter enclosure, m^2

L = length of the filter mesh, m

Thus the quantity $\frac{2FR_c dL}{\pi a}$ represents the fraction of particles captured in a differential length of the filter. Watson assumed that in a filter of randomly oriented filters, only two-thirds are oriented correctly to capture particles since fibers parallel to the magnetic field are ineffective. He also presented an argument showing that the wires in a differential thickness affect particles independently provided the packing density does not exceed a critical value. Critical values of F for selected values of R_c are shown in Table I.

TABLE I. MAXIMUM VALUES OF PACKING DENSITY FOR INDEPENDENT WIRE BEHAVIOR [1]

$\frac{V_m}{V_o}$	R_c (Case I)	Maximum F
400	9.45	0.0092
100	5.83	0.025
40	4.15	0.049
10	2.39	0.182
4	1.55	0.65

The collection efficiency expression was then integrated over the length of the filter to obtain

$$P = \frac{N_{out}}{N_{in}} = \exp\left[-\frac{4FR_cL}{3\pi a}\right] \qquad (3)$$

where P = penetration

N_{out} = number of particles out

N_{in} = number of particles in

The quantity (1-P) represents the fractional mass efficiency of the filter for any given size of particles, provided that all of the particles of that size are of equal mass density.

Other theories of HGMS have been advanced by Oberteuffer [3] and by Cummings, et al. [4]. Oberteuffer developed a force balance model without resorting directly to the equation of motion. His result is parametrically similar to Watson's but does not yield direct quantitative efficiency results. The major difference between Watson's theory and that of Cummings, et al. lies in the calculation of the magnetic force on the particle. In the present treatment (after Watson [1]) the magnetic field strength in the particle was assumed to be very nearly equal to the field strength outside the particle. In the strictest sense this implies that the potential energy of the particle is zero; however, it is easily shown that the result is the same as the more rigorous solution obtained by Cummings, et al. if the magnetic permeability of the fluid stream and the particle are nearly the same (as will be the case of paramagnetic particles in a gas stream). For the collection of ferromagnetic particles this simplifying assumption would be inaccurate.

RESULTS OF PREVIOUS HGMS APPLICATIONS

High gradient magnetic separation has been applied extensively to particle/liquid separations [3,5,6,7,8,9,10,11]. The primary commercial application has been in the removal of weakly magnetic color bodies less than 2 μm in diameter from kaolin clay. Pilot-scale and laboratory tests have been conducted on industrial waste process water from steel mills and electroplating operations, on nuclear reactor coolants, and on oils and hydraulic fluids (to remove wear particles). Preliminary experimentation has shown promise of application of HGMS to the desulfurization of coal and to the reduction of solids and dissolved orthophosphate from municipal waste. The latter application was preceded by seeding the waste stream with a magnetic material that became chemically or physically associated with the pollutant. Another recent novel experimental application was to the separation of red cells from whole blood.

Data have been reported from two studies in sufficient detail to allow comparison to the theoretical predictions of equation (3). In the first case Melville, et al. [7] studied the high gradient magnetic separation of red cells from whole blood. Red blood cells are flexible, biconcave, discoid bodies with well-defined dimensions of 8.5 ± 0.4 μm diameter and 2.3 ± 0.1 μm thickness. The filter contained 25 μm diameter cylindrical stainless steel wire. Flow velocities varied from 1.6×10^{-4} m/s to 5.8×10^{-4} m/s, and the magnetic field ranged from 0.6T to 2.4T. In all runs the parameter $\frac{\mu_o M_s}{2B_o}$ was less than 0.2. The experimental data and Case I theoretical prediction are shown in Figure 3. The equivalent spherical diameter assumed here for the purpose of modeling was 6.3 μm. For any given efficiency, the experimentally required value of $\frac{V_m}{V_o}$ is about 1 to 1.5 orders of magnitude higher than the theoretical value. Melville attributed this result to a postulated reduction in the actual value of R_c which occurred when the bare wire was covered with red blood cells within the first few seconds of filtration.

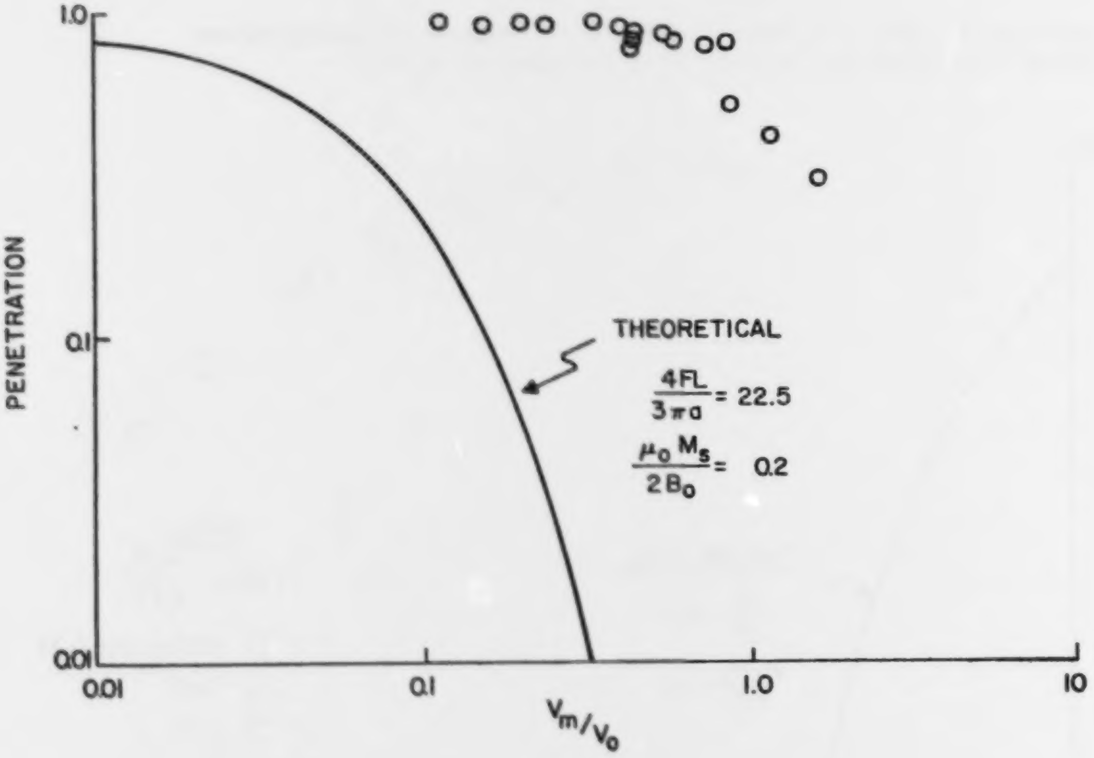

Figure 3. HGMS theory vs. experimental data: removal of red cells from whole blood.

Oberteuffer [8] reported data for the removal of 10 μm particles of cupric oxide from a slurry also containing diamagnetic aluminum oxide. Flow velocities ranged from 0.022 m/s to 0.22 m/s and the magnetic field ranged from 0.5T to 10T. Steel wool fiber packing was used. Luborsky and Drummond [12] reported additional details of the experiment and included in their analysis the assumption that the fibers could be approximated by an idealized-ribbon geometry (with width, S, and thickness, 2a). If only the small edges of this type fiber produce a significant magnetic force on the particles, then for a randomly packed filter only one-third of the fibers on the average would be oriented correctly. An argument analogous to that presented before then results in the theoretical expression

$$P = \exp\left[-\frac{FR_cL}{3S}\right] \tag{4}$$

The experimental data are shown in Figure 4. Theoretical penetrations calculated from equation (4) with Case I values of R_c fall

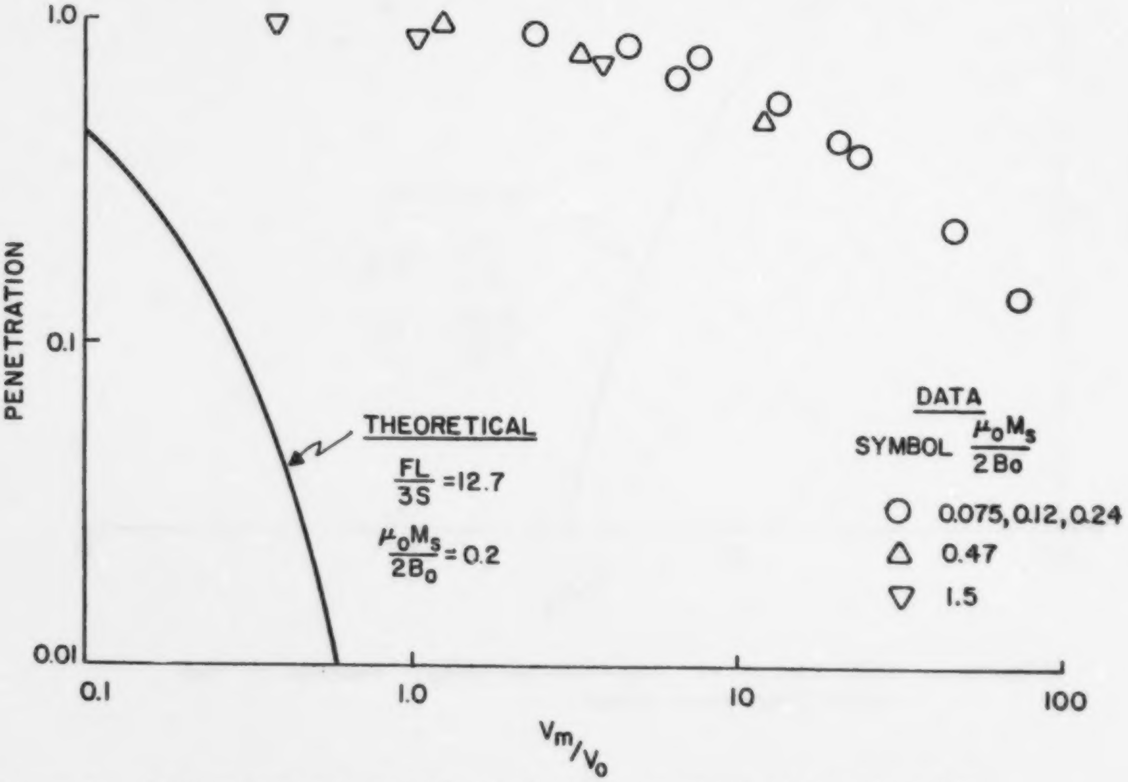

Figure 4. HGMS theory vs. experimental data: removal of cupric oxide particles from a water slurry.

roughly two orders of magnitude to the left of the data. Some of the data corresponding to $\frac{\mu_o M_s}{2B_o} > 0.2$ are plotted separately and tend to show a slightly enhanced collection. Luborsky and Drummond [6] were able to correlate these data with three modifications to Watson's model: (1) accounting for the effect on flow potential of particle buildup in an assumed geometric pattern (the field potential was assumed to be unaffected); (2) introduction of an empirically determined factor to account for the fraction of wires that are active in capturing particles (in addition to the one-third orientation factor); (3) introduction of an empirically determined factor to account for mechanical filtration of particles.

POTENTIAL APPLICATIONS OF HGMS IN PARTICULATE EMISSION CONTROL

Probably the most important parameter to be considered in the evaluation of potential application of HGMS is the particle susceptibility. Susceptibility of heterogeneous dust particles cannot be predicted accurately because it is strongly dependent on the distribution of magnetic species in the dust. The composition of industrial dust emissions obviously varies for different processes. Within a single process category the composition will vary over a more narrow range with respect to raw feed and operating variations. In Table II typical compositions of paramagnetic compounds are listed for several industrial processes. Blast furnace

TABLE II. POTENTIAL APPLICATIONS OF HGMS IN PARTICULATE EMISSION CONTROL

Industrial Process	References	Typical Dust Composition, mass basis
Blast Furnace	[13]	35-50% Fe, 12% FeO, 0.5-0.9% Mn
Basic Oxygen Furnace	[13, 14]	90% Fe_2O_3, 1.5% FeO, 4% Mn_3O_4
Open Hearth Furnace	[13, 14]	85-90% Fe_2O_3, 1-4 FeO, 0.5% MnO
Electric Arc Furnace	[13, 14]	20-55% Fe_2O_3, 4-10% FeO, 0.5% MnO
Silico-manganese Furnace	[13, 14]	4-7% FeO, 30-35% MnO
Ferro-manganese Furnace	[13, 15]	6% FeO, 34% MnO
Ferro-chrome Furnace	[13, 15]	7-11% FeO, 3% MnO, 29% total Cr as Cr_2O_3
Coal-fired Boiler	[14]	2-36% Fe_2O_3

dust, which has a high reduced iron content, should be relatively easy to collect. Since the dust would likely have residual magnetism, cleaning of the filter mesh could be a problem. Basic oxygen and electric-arc steelmaking furnaces are estimated to be prime candidates for HGMS control. Open hearth furnace dust has similar characteristics, but most existing open hearth furnaces are well-controlled, and the majority of new steelmaking installations will more likely be basic oxygen or electric-arc furnaces. In the ferroalloy industry, processes involving combinations of iron, manganese, and chromium have reasonable potential for HGMS application although the dust composition of these processes is highly variable. Coal-fired boilers are unlikely to be amenable to HGMS application, because the composition of magnetic species is relatively low.

Particle size also varies greatly in different types of dust. HGMS theory indicates that collection efficiency decreases with particle size, but experimentation will be required to determine the fractional efficiency for particular dusts. Gas temperature may also be an important factor. Particle magnetic susceptibility is reduced and gas viscosity is increased at increased temperatures. Both of these trends theoretically reduce HGMS efficiency, but the effect could be minor provided the efficiency at lower temperatures is sufficiently high.

DISCUSSION OF EXPERIMENTAL FEATURES

The bench-scale experiments reported in this paper were conducted with an 8.9 cm (3.5 in.) diameter HGMS system collecting dusts from a basic oxygen furnace. An operating temperature of 38°C (100°F) was used. Apparatus and procedures were discussed in detail in a previous paper [17]. The ranges of important operating parameters are summarized in Table III. Total mass concentrations on the order of 0.01 to 0.1 g/m^3 were produced by the BOF dust redispersion apparatus. Figure 5 shows the typical particle size distribution at the inlet of the magnetic separator as determined by Andersen impactor runs. The distribution is skewed more toward larger sizes than are typical primary emissions from a basic oxygen furnace, indicating that a significant mass percentage of the dust was still in an agglomerated form.

One long term run was made to study the variation in sampling results with time and the transient effects of mesh loading. The results are shown in Figure 6. The penetration (1-fractional collection efficiency) varied without any pattern for the first 8 hours of the test and then began to rise. After 10.5 hours the superficial gas velocity could no longer be maintained at its original value because of the increased pressure drop, and the run was stopped after 11.5 hours. At the end of the run the filter mesh was carefully removed and weighed. The clean weight was 120 g and a total of 81.8 g of dust had been collected. Assuming the total mass collection efficiency was constant over the 11.5 hour run (neglecting the small decrease in the last 4 hours), these

TABLE III. RANGES OF OPERATING PARAMETERS

Filter Mesh Number	4	5	6	7	8
Filter length, cm	35.6	20.3	20.3	20.3	20.3
Packing density	0.0088	0.0088	0.0174	0.0132	0.0132
Packing orientation	random	spiral	spiral	spiral	spiral
Magnetic field, T	0.094-0.308	0.094-0.308	0.094-0.214	0.094-0.214	0.094
Gas velocity, m/s	6.2-8.7	5.9-9.3	5.9-10.2	6.6-10.6	6.1-7.0
Pressure drop, kPa (in. H_2O)	1.2-2.0	0.7-1.3	1.4-3.2	1.2-2.9	1.2-1.4
	(4.6-7.8)	(2.8-5.3)	(5.4-12.8)	(4.7-11.5)	(4.6-5.4)

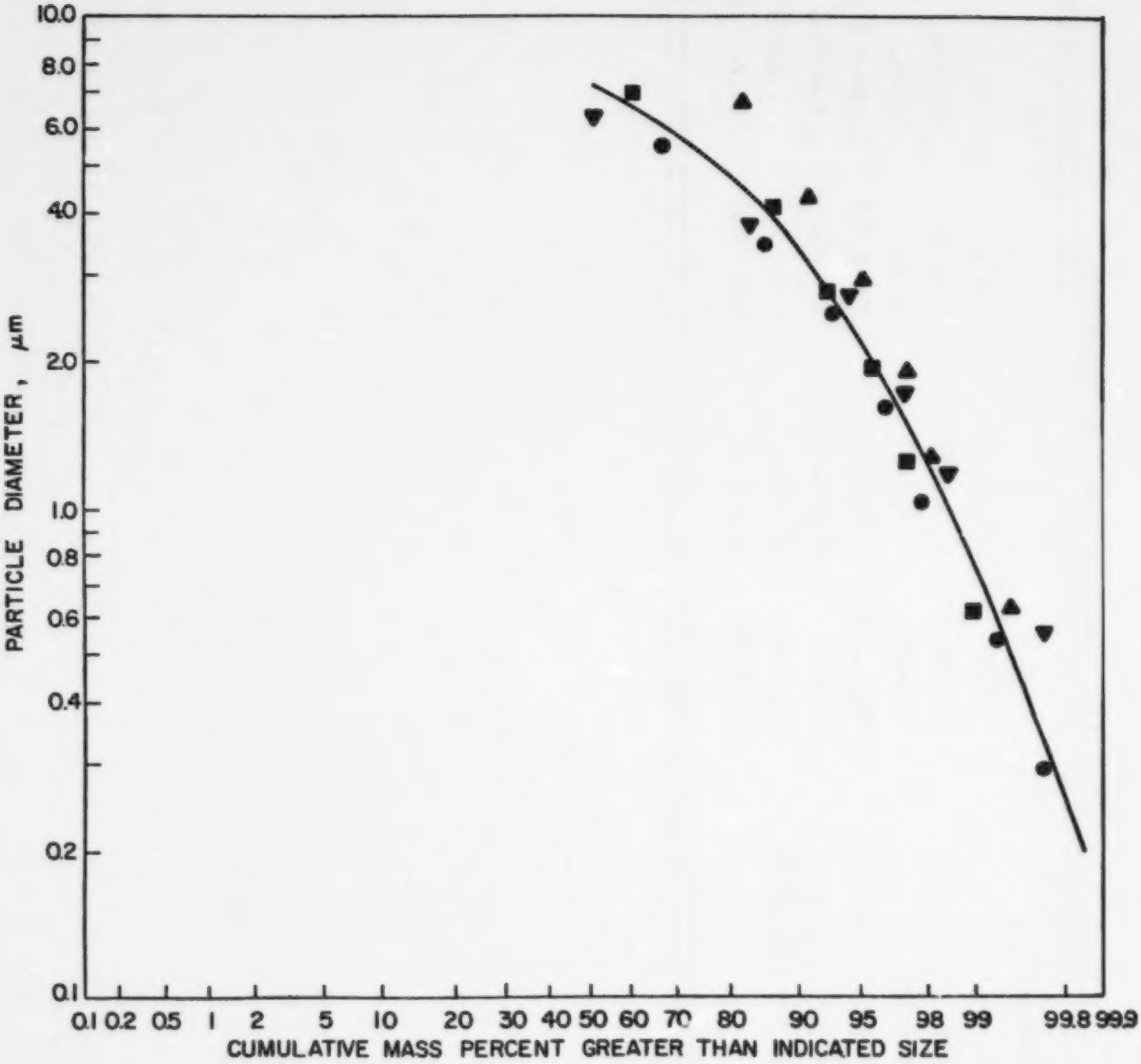

Figure 5. Size distribution of redispersed BOF dust.

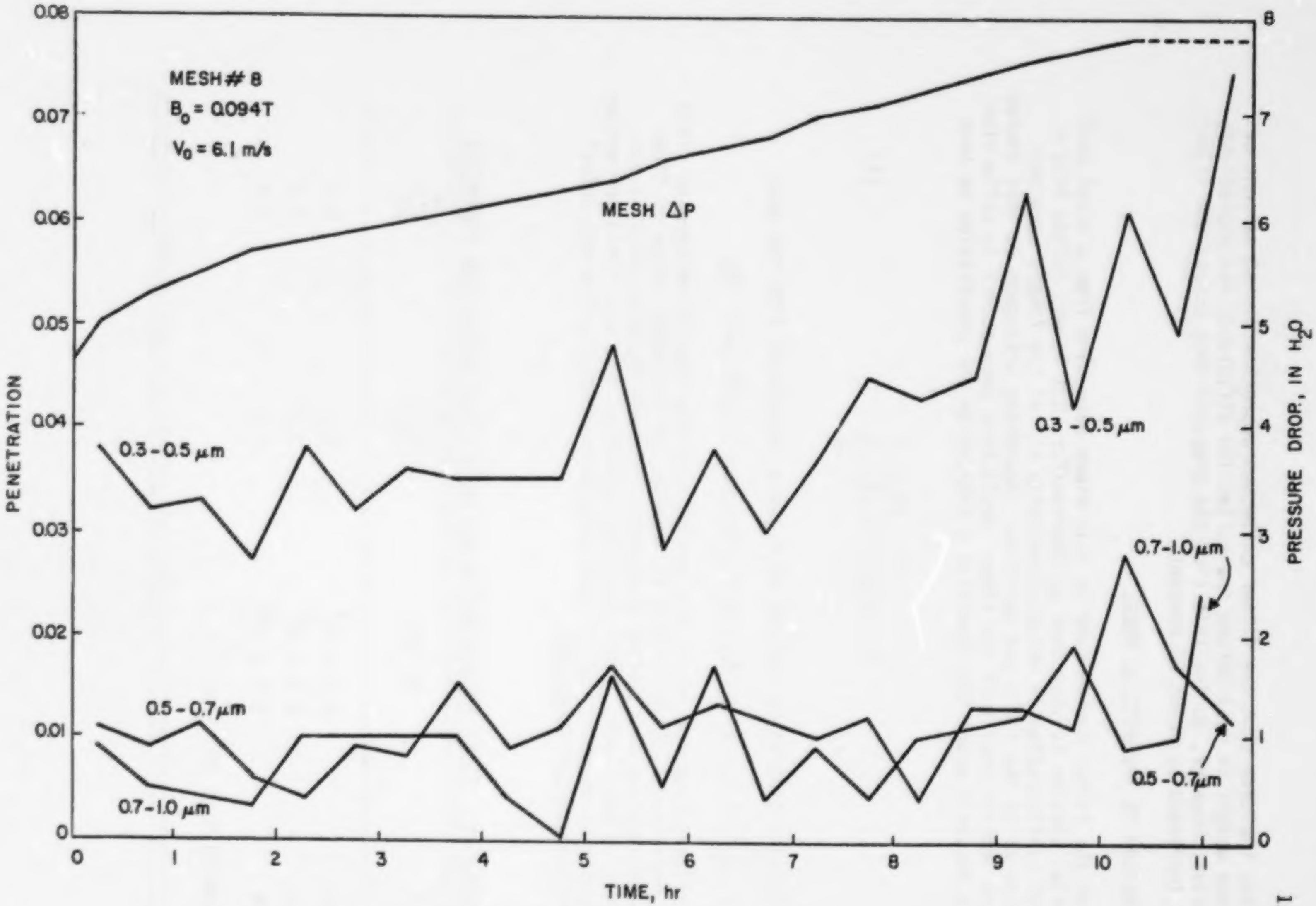

Figure 6. Transient loading effects and random variations in HGMS collection efficiency and pressure drop.

results indicate that the filter collected approximately 50 percent of its own weight in dust before the collection efficiency was appreciably affected. However, during this time the pressure drop across the filter also increased by about 50 percent.

COMPARISON TO THEORETICAL MODEL

All of the filter meshes used in this study were made from a steel wool material similar to that used by Oberteuffer, but it was rolled into a spiral configuration in which essentially all of the fibers were perpendicular to the field and gas flow. Reasoning analogous to that stated before implies that half the fibers would have been active in collection. Using Watson's model the theoretical expression for penetration becomes

$$P = \exp\left[-\frac{FR_cL}{2S}\right]. \qquad (5)$$

Appropriate theoretical values of R_c were determined from the case where $\frac{\mu_oM_s}{2B_o} = 2$. With the low fields used in this work the magnetization of the steel wool varied with the applied background field in a nonlinear fashion. Table IV reports the estimated value of fiber magnetization as calculated by Magnetic Engineering Associates, a consultant to the study. The calculated values of the near field correction parameter using estimates of both saturation magnetization and actual magnetization are tabulated.

TABLE IV. FIBER MAGNETIZATION AND NEAR FIELD CORRECTION PARAMETER

B_o, T	M, A/m	$\frac{\mu_o M}{2B_o}$	$\frac{\mu_oM_s}{2B_o}$
0.094	3.2×10^5	2.2	6.7
0.214	6.8×10^5	2.0	2.9
0.308	8.4×10^5	1.7	2.0

Estimated $M_s = 10^6$ A/m

At the time these experiments were conducted, equipment to measure the particle magnetization was not available. Based on the reported data of typical BOF dust composition the volume susceptibility was estimated to be 4×10^{-3} in dimensionless SI units. Recently, access to a Faraday apparatus was arranged and three BOF dust samples were processed with the results shown in Figure 7. The particulate is much more strongly magnetic than anticipated. At 0.094T, the apparent volume susceptibility with an assumed particle specific gravity of 3.5 is approximately 1.5 dimensionless SI units.

Typical experimental data from this study are presented in Figure 8 with comparison to the corresponding theoretical prediction. To calculate the abscissa value for the data, the estimated values $M = 3.2 \times 10^5$ A/m and $\chi = 1.5$ were utilized. In general this data set as well as the remaining data agree more closely with the theoretical prediction than the liquid system data of Melville and Oberteuffer. This feature is perhaps due to enhancement of collection efficiency associated with the near-field correction. The near field contribution is apparently important due to the combination of small particle size, low magnetic field and high fluid velocity. Table V summarizes the differences in the values of these variables for the three experiments. In view of the significant difference in the ranges of operating parameters as well as the extrapolation from liquid to gas systems, discrepancies between this work and the earlier experiments is not surprising.

Certain other features of the data should also be noted. Changes in the abscissa of Figure 8 represent changes in two parameters--gas velocity and particle size. Most of the scatter is believed to be associated with the measurement of particle size and concentration. Each data point represents the average of three or more upstream and downstream counts with an optical analyzer and sampling/dilution system. The individual penetration measurements for a given particle size during a given run varied by as much as 50 percent from the average. These variations were generally larger for 1 to 3 μm particles which appear as data clustered to the right of Figure 8. The ability to measure low penetration values for these large particles was also limited somewhat by low particle counts. Thus the apparent tendency for the data to level out toward larger $\frac{V_m}{V_o}$ values could reflect a sampling limitation rather than an actual physical trend. It should also be noted that the nominal particle diameters indicated by the optical analyzer are based on calibration with monodisperse latex spheres. Since the dust is known to have contained agglomerated particles (probably of various shapes), the relationship among the actual physical particle size and shape, effective particle size for interaction with the wires, and particle size indicated by the optical analyzer is not clearly defined.

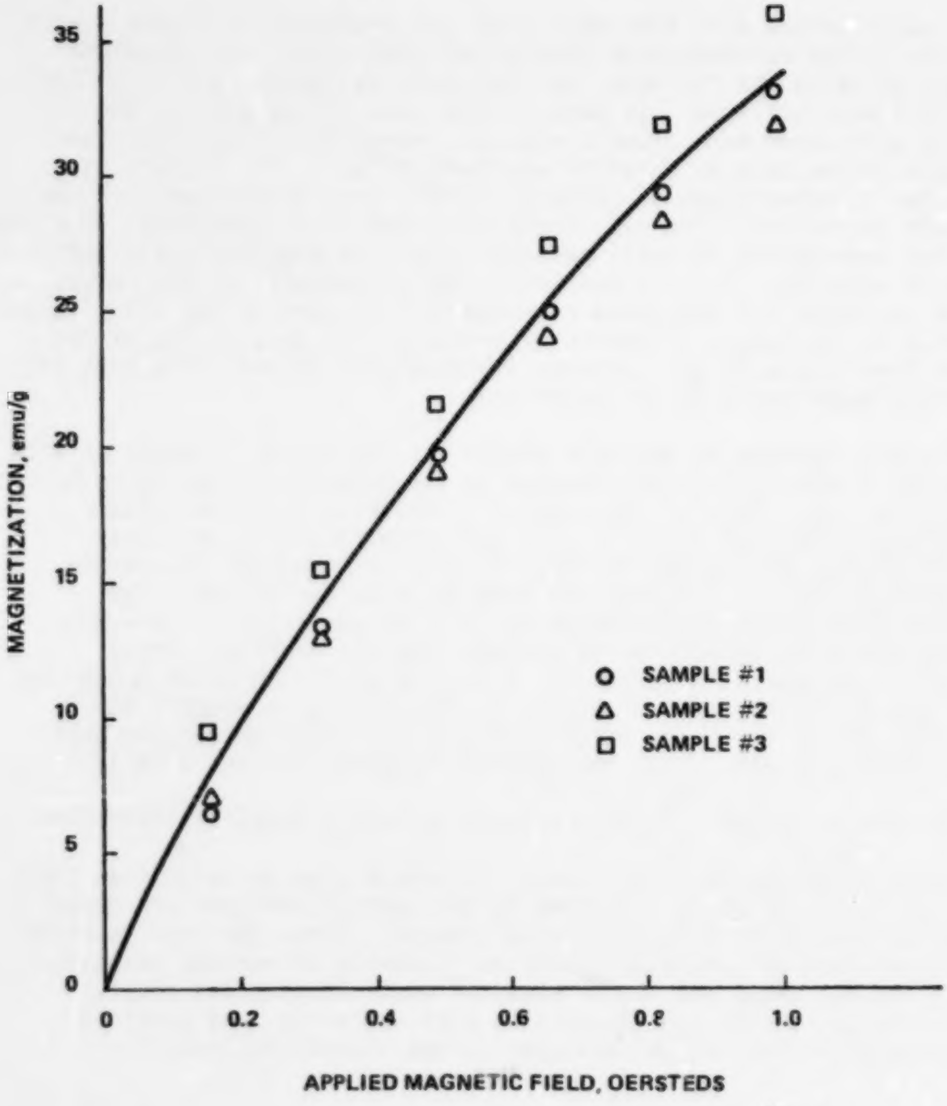

Figure 7. Experimental measurements of particle magnetization.

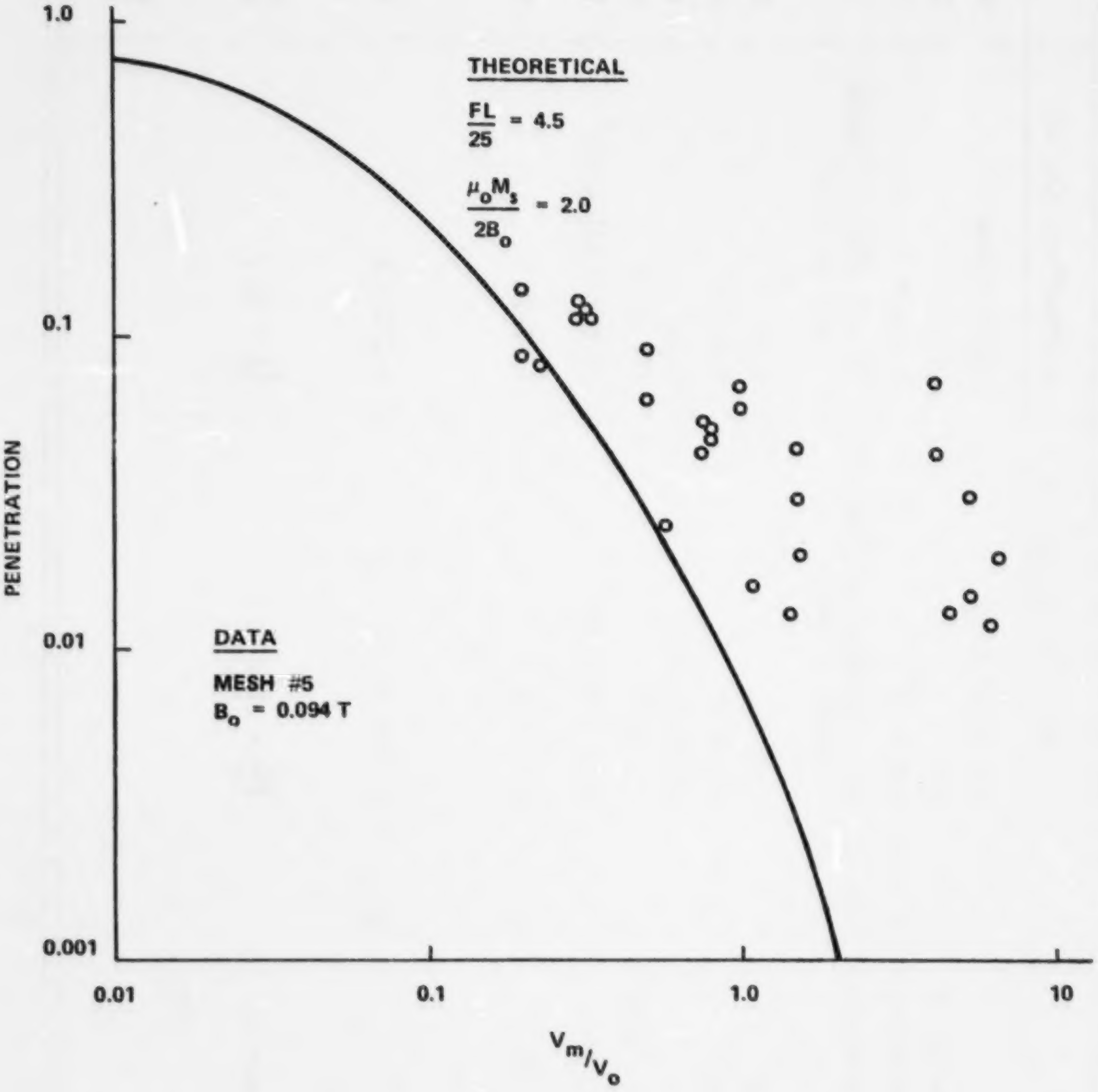

Figure 8. HGMS theory vs. experimental data: mesh No. 5, $B_o = 0.094T$.

TABLE V. SUMMARY OF DIFFERENCES IN THE PRESENT AND PREVIOUS WORK

Parameter	Melville, et al. [7]	Oberteuffer, [8, 12]	This Work
Particles	red blood cells	cupric oxide	BOF dust
Particles diameter, μm	6.3	10	0.3 - 3.0
Particle susceptibility, SI units	3.0×10^{-6}	2.4×10^{-4}	1.5
Fiber type, geometry and size	25 μm diameter cylindrical stainless steel	20 μm x 200 μm ribbons stainless steel	20 μm x 200 μm ribbons stainless steel
Packing density	0.02	0.05	0.0088-0.0174
Magnetic field, T	0.6-2.4T	0.5-10T	0.094-0.308T
Fluid	diluted blood	water	air
Fluid viscosity, Ns/m^2	1×10^{-3}	1×10^{-3}	2×10^{-5}
Fluid velocity, m/s	$(1.6-5.8) \times 10^{-4}$	$(0.022-0.22) \times 10^{-2}$	5.9-10.6
$\frac{V_m}{V_o}$	0.11-1.6	0.37-73	0.17-30
$\frac{\mu_o M_s}{2B_o}$	0.04-0.17	0.075-1.5	2.0-6.7
Packing parameter	$\frac{4FL}{3\pi a} = 22.4$	$\frac{FL}{3S} = 12.7$	$\frac{FL}{3S} = 5.2$ $\frac{FL}{2S} = 4.5-8.8$

Attempts to compare the effects of individual parameters to the theoretical predictions were thwarted primarily by the relatively narrow range of the individual parameter variations and the scatter in the experimental data. The nonlinear relationship of R_c and $\frac{V_m}{V_o}$ also complicates individual parameter analysis. The strongest evidence of some physical mechanism not accounted for in Watson's theory was seen in analysis of penetration-vs-velocity data with other parameters constant. The theory predicts for $\frac{V_m}{V_o} < 0.2$:

$$P \simeq \exp\left[-\frac{k}{V_o}\right] \tag{6}$$

where k is a constant; i.e., as velocity increases, penetration increases. In several cases the data tend to show penetration decreasing with increasing velocity or perhaps a minimum penetration at a mid-range velocity. Figure 9 shows one of these cases. This behavior conceivably reflects a complicated interaction of high gradient magnetic separation, inertial impaction (which is neglected in Watson's development), and reentrainment.

With respect to the magnitude of observed collection efficiencies, the data in Figure 8 indicate penetrations on the order of 0.01 for particles in the 1 to 3 μm range and 0.15 for particles in the 0.3 to 0.5 μm range with a relatively low applied magnetic field. In other runs when the magnetic field or the filter mesh density was increased penetration of the 0.3 to 0.5 μm particles was easily reduced to 0.05 and below.

CONTINUING WORK

Basically the objective of this research is to demonstrate on a small scale that high gradient magnetic separation may be an effective and economic technique for the control of fine particle emissions from selected industrial sources.

To satisfy this objective, experiments are continuing on a pilot-scale HGMS apparatus of approximately 0.7 m^3/s (1500 CFM) capacity. At least two levels of the following parameters will be studied: superficial gas velocity, gas temperature, magnetic flux density, particulate source, and particulate size.

Particle collection efficiency will be correlated as a function of operating parameters either by an extension of Watson's theoretical model or if necessary, by a semi-empirical approach. The primary purpose of

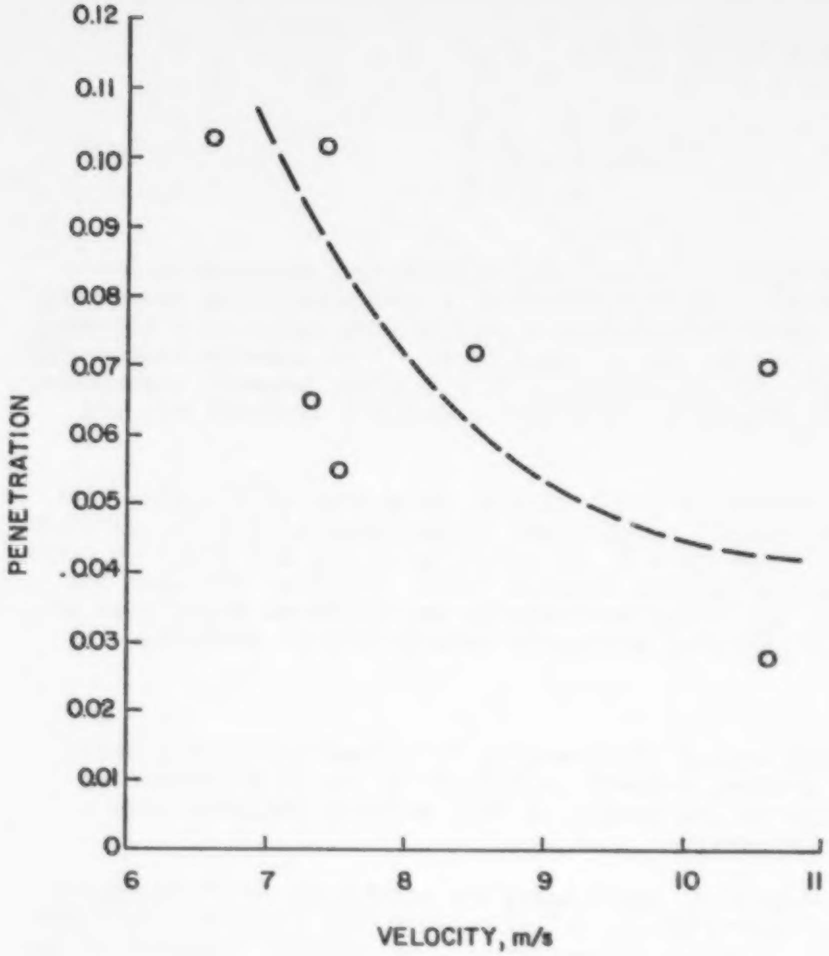

Figure 9. Anomolous effect of velocity on penetration: mesh No. 7, B_o = 0.094T, 0.3-0.5 μm.

the correlation will be to develop a reliable basis for designing a larger scale device with a high probability of success. A secondary goal, which is desirable but may prove to be beyond the scope of the present research, is to develop a model of sufficient completeness and accuracy to allow elementary optimization techniques to be incorporated in the scale-up design (i.e., to provide a basis for determining the incremental cost effect associated with adjustment of the major operating parameters at constant collection efficiency so that each of these parameters can be set at its optimum value.) Although Watson's model has been useful so far, it may not be sufficient without modification for at least three reasons:

1) The experimental data indicate a velocity effect that is not predicted by the model;
2) At low fields the fibers are not magnetically saturated as the model assumes;
3) The particles are ferromagnetic rather than paramagnetic as the model assumes.

CONCLUSIONS

The following conclusions have been drawn from the preliminary experiments and theoretical analysis conducted to date:

(1) Fundamental theory predicts the potential success of the concept.

(2) Preliminary experiments support the potential capability of at least 90 percent removal of all submicron particulate from BOF furnaces. Collection of submicron particulate might actually be enhanced at low magnetic fields.

(3) No inherent health and safety hazards or any other unfavorable environmental impacts have been associated with HGMS application.

(4) No prohibitive design and reliability problems have been identified.

(5) HGMS is potentially applicable to several sources of fine particulate; retrofit capability should be enhanced by the high gas velocities demonstrated in preliminary experiments.

ACKNOWLEDGMENTS

The authors are indebted to Dr. Larry K. Monteith of North Carolina State University for major assistance in the mathematical model interpretation.

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WRITTEN DISCUSSION

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Because this technique is based on intrinsic magnetic susceptability its application must be limited to such materials that have magnetic properties. On a practical level the separation of fine particulates in a large volume gas stream may only be possible with materials of high magnetic susceptability.

The experimental results of the Research Triangle Institute (RTI) do support the validity of the Watson model. It is especially interesting to note that by using the measured (rather than assumed) particle susceptability a much better fit is found to the theoretical model.

During the aerodynamic and magnetic field characteristics of the spaces between the stainless steel fibers is perhaps the most difficult analytical task. The usual model assumes a uniform spacing proportional to the matrix density and fiber size. This ignores the actual situation where there is a large random spacial variation, from touching to many fiber diameters separation. It is probable that the ratio of widely spaced to closely spaced fibers is decreased with increasing matrix density. This will have a profound effect on the capture of smaller size particles where the viscous forces are proportionately larger than for large particles. Investigation into the specific relationship of capture radius with respect to particle velocity, the geometric spacing and magnetic field distribution of adjacent fibers is needed.

Predicting the performance of a high gradient mangetic separator for air carried particulates from the results of water mixed experiments has not been successful. In the Stokes relationship there is orders of magnitude difference in the viscosity and density ratio between air and water. Thus any scaling would be quite sensitive to the actual experimental conditions.

A steel wool matrix would probably show a poorer correlation than an expanded metal one because of greater displacement effect of high density water on individual fibers. The literature contains some analyses of particle trajectories in a low density medium. However none has considered local density variations in a high velocity air stream passing through a fiber array and how this may affect the path of very small particles in the superimposed magnetic field gradients.

The high magnetic gradient filtration of air carried ferric oxide particulate from steel making processes ia a very appropriate application being investigated by RTI. It is also being evaluated by other groups for desulphurization of an air powdered coal mixture.

Another potential application lies in the filtration of petroleum gas streams containing finely divided iron and sulphur compounds. A rather unusual problem is found in high temperature utility steam turbines. Where iron oxide particles exfoliate from the pipe walls and are carried with the stream into the turbine causing blade erosion or in the extreme, blade rupture. It is possible to concentrate the particles with inertial and magnetic gradients and separate this from the main steam flow into a high gradient magnetic separator.

In all of these applications the capability of operating up to 500C is a distinct advantage of the Magnetic Precipitator. (This designation has been suggested by Henry Kolm and seems very appropriate to describe the action of high magnetic gradient devices for air/particulate mixtures.) Of course the saturation field for the matrix material will decrease with increasing temperature, (to zero at the curie point) which could result in a requirement to increase the magnetizing field to compensate.

The design of magnetic precipitator for large volume process rates most conveniently takes the form of several large solenoids in parallel. The collected magnetics are cyclically flushed out of each unit in turn by turning off the magnet. A conventional bag house or spray collector is needed to receive the blow down. Because the buildup time for the magnetics is long for the usual low loading rates, flushing may be needed at 5 or even 20 minute intervals and experiment has shown flushing is complete in less than 5 seconds at process flow rates. Because of this, the flush may be dumped in a large volume receiving tank and the particulate recovery filter operated at perhaps 5% of the system flow rate. Otherwise the flushing filter requirements would dominate the whole installation which is completely unacceptable. The size and number of solenoid magnetic matrix units would depend on the particular application, with consideration for flow rate, operating cost and capital costs.

The alternative designs use a moving matrix, either rotating or oscillating, where collection of the particulates in the magnetic field is continually passes through a field free region for flushing. The two principle differences compared to the solenoid geometry are decreased magnetic efficiency and the need for moving seals to maintain the pressure difference across the matrix. These requirements must be evaluated against the need to valve or direct a large volume air flow from flushing to a receiver for each solenoid. Treating the flushed mags is the same for each design.

REVIEW OF EXPERIMENTAL RESULTS

The experimental results on BOF dust obtained by RTI in their laboratory work using a SALS HGMS system clearly show a velocity dependent penetration fields up to 3 kilogauss. The effect is most pronounced for particles smaller than about 0.7 μm and is field dependent with a minimum in the 1400 to 1700 ft/min velocity range. For particles of 1 to 2 μ (and larger) diameter magnetic saturation takes place at 2 kilogauss with almost 100% filtration efficiency.

A logical explanation for this velocity dependency suggests that at lower flow rates the particle flow is laminar and stream lines do not come close enough to the fibers for capture in the magnetic gradient. At the highest flow rates the fluid drag is stronger than the magnetic attraction and particles are reentrained.

A long term run to measure the maximum loading capacity of the matrix resulted in confirming the usual assumption that the accumulation can equal the matrix weight. Insufficient data was taken on the effect of matrix length to report but penetration appears to be inversely linear with length.

There is a strong inverse dependency of matrix density and penetration for particles of less than 0.5μ the minimum penetration decreases from 8 to 2% for the smaller particles to the reduction of average and of the maximum distance between matrix fibers, but is obtained at a two fold increase in the pressure drop across the matrix. Since pumping power is about 10 times more costly than magnet power it appears there will be a trade off between a maximum achievable reduction in the smaller particulate effluent and an economically acceptable performance.

One aspect of experiments with air entrained particulates is finding a satisfactory technique for introducing properly dispersed material. So far, the use of a fluidized bed has been used to break up agglomerations of particles, but the maximum size which can be dispersed in near 30 μm and the total mass flow which can be entrained is perhaps 50 times smaller than desired loading per cubic foot of air. Other particulate dispersal methods should be developed to give a much broader control over experimental particle loading rates.

OPEN DISCUSSION

Kelland: When the particulate is on the side of the fiber, the gradient is in the reverse direction. Not only does the viscous force take the particle off but the magnetic force tends to push it away from the fiber also. However, there were some experiments done 5 or 6 years ago at MIT by Dr. John Oberteuffer in which expanded metal matrices were used in three different orientations to measure collection efficiency. Almost no difference in collection efficiency was found. How that translates to the practical devices of today I am not sure.

Gooding: Charlie pointed this out also; that his fibers were not round but were usually more flattened. The orientation of the flatness may have an effect on separation efficiency.

Friedlander: I disagree with that and have pictures to prove otherwise! The capture effect change is not true because the magnetic force tries to hold them there. What happens at higher velocities, however, I do not know.

Olson: Mr. Gooding has followed Mr. Watson's analysis very well, however, if one considers a particle trajectory model a drier separation occurs. In particular, the conditions that Gooding and Drehmel considered (Stoke's number between .1 and 30) is a region on the low side of transition; from a noninertial region where the particle inertia is not important to a region where the particle inertia is very important. So, in response, as Gooding suggested, there will be inertial impaction in filters over much of the velocity and particle sizes he considered.

The same would also be expected of the more conventional high gradient magnetic separator Robin Oder considered. In addition, in the Gooding-Drehmel work, the Stokes drag turns out not to be a very good approximation for the viscous forces on the particle. They are in the transition region where the Stokes theory for the flow around a particle breaks down. So there will perhaps be a different dependence upon the relative particle fluid velocity rather than a linear one.

Gooding: I would just say that I agree with what you said. This is something we have put some thinking into. But we have not gone through to try to improve the model.

Friedlander: It seems to me that the performance goes in the opposite direction than what Maxwell said a little while ago about much worse performance at higher velocities. If I read your graph correctly, improved performance comes with higher velocities.

Gooding: This is a single component system where you are not trying to separate two different solid particulates in an airstream. It is a much simpler problem than the problem where you have two components that you are trying to separate. I would not compare them at all.

Friedlander: Without taking away from Watson, I think one has to say that Zebel did the basic thing. We should give credit to Zebel.

Oder: I think we need a knock-down drag-out session just on magnetic separation. Now you know why I chose the example I did, I knew the magnetic component you would be looking at. I jimmied up the Hazen process to come out with moments in the same ballpark. We are not that far apart and your flow velocities are of the order of 1000 ft/min. If you can convert coal minerals to a very strong magnetic component, you get some indirect proof that it should work.

Gooding: I think your velocity dependence looks like inertial deposition. I think there are two things going on, inertial deposition and reentrainment. I have played with the mathematics a little bit. We may find that there is a minimum or optimum velocity.

Oder: Did you ever run with your magnet turned off?

Gooding: Yes. The figure shows no field, and you see the larger particles are getting as high as 50% collected. There is a fair amount of uncertainty in this data. This is one of the reasons we have not tried to do any more sophisticated modeling. Referring back to this slide, we had indications that some of the larger particles were agglomerated as they impact on the wire, which would mean that the zero field numbers are suspect. We have improved that aspect of our experimental apparatus by building a scaled-up version. We are also taking data as a baseline at zero field.

When you are dealing with a ferro-magnetic particle, another force that you really should consider is one that you do not normally calculate in the analysis. If you have a ferro-magnetic particle, particularly a very small one and it gets impacted on a ferro-magnetic surface which is larger, the image force due to the moment in the ferro-magnetic wall surface should be considered. In other words, the particle itself is a source of magnetic field gradient which makes it stick to the wall. This operates only at distances comparable to the diameter of the particle. It is something which should be taken into account when you are dealing with ferro-magnetic particles. This may have some relation to what you seek in your zero field measurements. Because, even though you turn your field off, if this is steel wool or something like that, it does have a small amount of residual magnetism. It will pick up ferro-magnetic particles. I believe you are correct. In our present experiments we have designed it so that all our zero magnetic field data is taken before we magnetize the matrix.

RESEARCH NEEDS AND OPPORTUNITIES IN HIGH GRADIENT MAGNETIC SEPARATION OF PARTICULATE-GAS SYSTEMS

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ABSTRACT

This paper suggests the major needs and opportunities in high gradient magnetic separation (HGMS) of particulate-gas systems according to three categories: (1) process and engineering research oriented towards effective applications of existing devices; (2) fundamental research aimed at developing a better understanding of the mechanisms and principles of the separation; and (3) scientific and engineering development of new separation methods and devices.

INTRODUCTION

High gradient magnetic separation (HGMS) is a new technique that provides a practical means for separating micron-size, feebly magnetic materials on a large scale and at much faster flow rates than are possible in ordinary filtration. The technique is also applicable to separating nonmagnetic materials which can be made to associate with magnetic seeding materials. HGMS has already been applied industrially for the wet cleaning of feebly magnetic contaminants from kaolin clay with very low costs and outstanding technical performance {15, 16, 24, 28, 29, 34-36, 38}. It is also proving applicable to solving many important problems related to energy production (e.g., coal desulfurization), environmental control (e.g., particulate collection and wastewater treatment), resource recovery (e.g., mineral beneficiation), and other minerals and chemical processing problems. The significance of the HGMS, and its latest engineering and commercial developments have been adequately discussed elsewhere {16, 28, 29, 35, 36, 38}.

Because of the necessary costs for drying and dewatering the products in most wet applications, it is generally believed that the most economical application of the HGMS is for the dry separation of many particulate-gas systems, especially those particulates of less than 2 to 3 microns in diameter. Existing experimental data have already demonstrated the technical feasibility of utilizing the HGMS for the collection of the basic oxygen furnace (BOF) dust {10}, and for the removal of pyritic sulfur and ash from dry pulverized coal {32}. These reported results and other unpublished data known to the authors have suggested that the HGMS appears to hold much promise as an effective alternative to particulate-gas separation, and further research and development work is clearly needed.

The purpose of this paper is to briefly suggest the major needs and opportunities in the HGMS of particulate-gas systems. For convenience, the discussion will be divided into three categories: (1) process and engineering research oriented towards applications of existing devices; (2) fundamental research aimed at developing a better understanding of the mechanisms and principles of the separation; and (3) scientific and engineering development of new separation methods and devices.

APPLICATIONS-ORIENTED PROCESS AND ENGINEERING RESEARCH

Applications-oriented process and engineering research is needed for demonstrating that the HGMS is an effective and economical method for separating particulate-gas systems. This type of research may include the development of magnetic process for particulate-gas separation which has been previously considered to be impossible or uneconomical; and the adaption of the magnetic approach to improve the performance of existing particulate-gas separation devices. Obviously, there are numerous areas or opportunities for applying the existing HGMS devices to particulate-gas separation. Three practical examples of such applications related to energy production, environmental control and resource recovery can be cited here to illustrate the broad range of research areas or opportunities available.

Among the many areas in which the HGMS may have potential applications to particulate-gas separation, the magnetic removal of pyritic sulfur and ash from dry pulverized coal is believed to be particularly important and promising {19, 21, 24-27, 32, 37, 38}. Substantial reduction in both pyritic sulfur and ash contents in the HGMS of several Illinois Basin coals has already been reported in a recent study by Murray {32}. An interesting observation from this study is that the combined use of several finely divided filamentary matrix materials such as steel wool, steel filings, etc., can significantly enhance the magnetic removal of both pyritic sulfur and ash. This observation confirms the previously recognized importance of the matrix characteristics in providing strong high-gradient magnetic fields {16, 29, 34, 35, 38}. Clearly, research and development work in optimum matrix design and composition will continue to contribute significantly to advancing the state of the HGMS technology, especially in applying to particulate-gas separation. Another important area which has not received much attention is the optimum use of the latest fluidization technology in conjunction with the high gradient magnetic desulfurization of dry pulverized coal. Identifying the problems associated with fluidizing pulverized coal, determining the conditions for achieving good fluidization, and characterizing the performance of the separation are only some of the work needed.

Because of the national interest in increasing the utilization of coal, more financial support for process and engineering research related to the magnetic desulfurization of dry pulverized coal is being provided by several governmental and funding agencies. However, the results from a recent survey of the supported research projects on the subject conducted by the authors have suggested that the current level of research support does not seem to reflect the proper appreciation of the significance of the HGMS, and the potential and economics of the magnetic cleaning of dry pulverized coal. The latter has been adequately discussed in the literature {21,29, 32, 37, 38}. Here, it is sufficient to mention that the new HGMS technology has provided a possible commercial method for economically separating the sulfur-bearing and ash-forming minerals from dry pulverized coal on a large scale. It has been estimated that a total of 100 million short tons of U. S. coals per year, low enough in organic sulfur, can be magnetically cleaned as an environmentally acceptable, low-sulfur low-ash fuel. This amounts to over 17% of the total U. S. production per year {25-27}. Further, recent analyses have clearly indicated that the costs of high gradient magnetic desulfurization, especially of dry pulverized coal, appear to be attractive when compared to those of other pyrite removal processes currently under active developments {23, 37}. All of these aspects will suggest that increased research support for the work on the magnetic desulfurization of dry pulverized coal is most likely to produce the greatest dividends.

The recent work by Drehmel and Gooding {10} has illustrated the potential of applying the HGMS to particulate-gas separation in the important area of environmental control. In particular, the technical feasibility of utilizing the HGMS for the collection of the basic oxygen furnace (BOF) dust has been demonstrated. An important observation from this work is the dramatic improvement of particulate collection efficiency resulted from applying a relatively low-intensity magnetic field. Total mass collection efficiencies of 99.9 percent or better were found with a 3.08 kilogauss field, which compared quite favorably with the typical efficiencies of 50 percent or less observed without applying the magnetic field. However, it was also found from this study that the existing theory of HGMS failed to interpret quantitatively the effect of some separation variables such as particle size, field intensity, etc. Despite the preliminary nature of this investigation, it is apparent that the HGMS holds much promise of application to the important problem of fine particulate emission control, especially in the iron, steel, ferroalloy, and other related industries. This follows because a majority of the particulate matters in these industries are sufficiently magnetic to be separated by applying a high gradient magnetic field. The primary need at this point is to identify clearly several potential areas of application in these industries by a close interdisciplinary cooperation between the particulate control practitioners and the magnetic separation technologists. Further, it is anticipated that many technical problems will arise in applying the existing HGMS devices to some of the selected areas. This necessitates some proper process and engineering modifications of the current HGMS technology. For instance, effective means must be developed to allow the use of the present HGMS devices in separating the high-temperature particulate-gas mixtures such as these off gases from steelmaking furnaces; and to prevent the chemically corrosive components such as chlorine existing in some particulate-gas mixtures from corroding the separator matrix. The major objective of such process and engineering work should be, of course, on demonstrating that the desired particulate-gas separation can be carried out effectively by the HGMS at a low enough cost to be economically attractive when compared to those of the competing fine particulate control technologies {1, 11, 12, 31 40}.

Applications-oriented research has also been initiated in this laboratory and elsewhere on utilizing the HGMS to recover the resuable components from the fine airborne dusts emitted from industrial operations. This type of research attempts to extend the potential applicability of the HGMS for separating particulate-gas systems in environmental control to an increasingly important area of resource recovery. As an illustration, it is known that fly ash from coal-fired power plants contains a large fraction of reusable, fine magnetic particulates such as magnetites. Existing analysis has already suggested that they can be economically recovered by the HGMS, thus reducing the load on the more expensive electrostatic precipitators {29}. Preliminary work in this laboratory and unpublished prior work in the coal mining industry have also indicated that the coal ash derived magnetite performs better in the heavy-media coal cleaning process than does mined magnetite {26}. In view of the large amount of fly ash generated from coal-fired power plants in the U.S., and the increasing cost of the commercial-grade mined magnetite being used in coal preparation, further research toward the development of an economical ash-derived-magnetite coal cleaning process is clearly justified. It is also apparent that there are many other areas or opportunities related to resource recovery, in which the use of HGMS for separating particulate-gas systems plays a major role; and they are too diverse to be surveyed here.

Regardless of the area of application, process and engineering research oriented towards the effective and economical utilization of the existing HGMS devices for particulate-gas separation should normally include the following steps: (1) examination of the evidence of significant magnetic removal of particulates from their carrier gas; (2) study of the quantitative effects of operating conditions such as field intensity, residence time, etc. on the grade and recovery of the separation; (3) study of the separator capacity and scale-up; (4) conceptual process design and cost estimation; and (5) comparison of costs and performance with those of the competing technologies. Here, the recent experience in carrying out most of these steps in the development of HGMS processes for coal desulfurization and kaolin beneficiation reported in the literature {21-28, 33, 34, 37-39} is believed to be helpful to those involved in the process and engineering research on the HGMS of particulate-gas systems.

FUNDAMENTAL RESEARCH

The key to successful utilization of HGMS devices for particulate-gas separation is the clear understanding of the mechanisms and principles involved in relation to specific applications. Recent experience gained in the pilot-scale and commercial applications of HGMS to coal desulfurization and kaolin beneficiation {21, 22, 27, 33, 37-39} has demonstrated the importance of utilizing quantitative separation models in conjunction with the separation testing. In particular, the successful experimental verification of two simplified mathematical models due to Bean {5}, and Oder and Price {39} in such reported applications has allowed one to quantitatively predict the trade-off of some operating parameters such as field intensity, residence time, etc., so as to optimize the magnetic desulfurization of coal and beneficiation of kaolin {22, 39}. None of these two simplified models, however, could satisfactorily predict the phenomena of mechanical entrapment and matrix loading (concentration breakthrough) observed experimentally.

Similar research related to the above fundamental aspects of HGMS applied to particulate-gas separation is, unfortunately, being pursued much less vigorously. This may be exemplified by the considerable uncertainties in the theoretical understanding of the quantitative effects of separation variables in the recent study of high gradient magnetic particulate collection by Drehmel and Gooding {10}. Clearly, in order to permit a more scientific approach to future applications of HGMS to particulate-gas systems, much work is needed to develop the necessary conceptual understanding and the proper modeling framework. From a detailed review of the recent development and current status of the quantitative modeling of HGMS described elsewhere {28}, it may be suggested that such a modeling framework should at least take into account of the phenomena of mechanical entrapment and matrix loading, as well as the characteristics of the particulates to be separated. A proper approach accounting for such factors also necessitates the conventional matrix-type HGMS device to be modeled as a magnetic adsorption column as suggested by Kaiser, et al. {17}. An example of a typical mathematical model based on this approach can be found from the work in this laboratory {28, 33}, in which the successful application of the modeling results to quantitatively predict the experimental data from coal desulfurization is also demonstrated. It is believed that the same modeling framework can be properly extended for predicting the performance of HGMS of particulate-gas systems.

The success of the preceding attempts in developing quantitative models to satisfactorily predict the overall performance of the separation such as grade, recovery, etc., can be greatly enhanced by simultaneously conducting experimental studies on some synthetic and well-defined particulate-gas systems. Such experiments may be carried out, for example, by using air-entrained monodispersed particles with uniformly spaced wires as collecting fibers in the separator matrix. Here, a number of similar experimental studies of the HGMS of particles in water slurries reported in the literature {6-9, 14, 18} may serve as good references. The experimental results thus obtained should provide some helpful insight into the mechanisms of particle capture and buildup, etc., in particulate-gas systems.

EQUIPMENT DESIGN AND DEVELOPMENT

Research in this category includes the improved design and scale-up of the existing HGMS devices, as well as the development of devices utilizing new separation concepts. The importance of this type of research can be readily recognized by noting the significant effort currently being devoted to developing new separation methods and devices, including those suitable for dry separation of particulate-gas systems {3, 4, 13, 15, 16, 36, 42}. Further, the commercial breakthrough in kaolin beneficiation resulted from introducing a new level of HGMS unit in 1969 has also made it clear that expanded research activity in equipment design and development will offer the greatest opportunities in advancing the technology of HGMS of particulate-gas systems.

A number of research needs and opportunities in equipment design and development can be briefly suggested by examing the status of several recent technological innovations in HGMS, and their potential applicability in particulate-gas separation. First, it has been generally agreed that improved or new design, and scale-up of the original production prototype HGMS unit have been made during the past few years; and stainless steel wool is widely accepted as a common magnetic matrix. What has been largely neglected here is the systematic research aimed at developing alternative matrix collectors, especially for use in particulate-gas separation. Next, for certain separation problems in which the feed stream contains a large fraction of magnetic particles, technical and economical constraints often necessitate the use of continuous HGMS devices to increase the separation throughput. For example, both moving-matrix Carousel separator {36} and open-gradient stream splitter {3, 4} are now being developed for such separations. Conceptually speaking, the stream splitter is especially attractive because it is a truly continuous device in which the magnetic particles, rather than being collected on a matrix, are deflected to one side of the splitter by the magnetic field; whereas the Carousel separator contains moving parts, and sometimes does not operate in a truly continuous fashion. Because of the lack of the gradient-producing collection matrix in the stream splitter, it has not been proven practical for separations of fine paramagnetic particles at present. Thus, it has been suggested that stream splitting devices are best used in conjunction with the high-intensity magnetic fields generated by superconducting magnets {13}, which appears to hold much promise of application to separating particulate-gas systems. Here, the open space existing in the superconducting stream splitting device allows for the use of the latest fluidization technology to enhance the particulate-gas separation. Clearly, further work is needed to quantitively characterize such an approach. Alternatively, there is also a need to develop a high-intensity, high-gradient stream splitting device, without using the superconducting magnet. The recent work by Allen and Iannicelli {2} illustrates one promising approach to this development, and this new design has yet to be tested in actual separations. Finally, perhaps the most important need in equipment design and development is an increase in research support, as significant advances can be made in particulate-gas separation with the development of the new generation of HGMS technology.

ACKNOWLEDGEMENT

The authors wish to gratefully acknowledge the financial support for their work on magnetic separation provided by the National Science Foundation, the Gulf Oil Foundation, the Board of Trustees and the Engineering Experiment Station of Auburn University, and the U. S. Energy Research and Development Administration- Oak Ridge National Laboratory.

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WRITTEN DISCUSSION

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Most of the work done on applying High Gradient Magnetic Separation (HGMS) to particulate-gas systems apparently has been in the area of coal desulfurization. For the purpose of this paper as stated in the title it would be worthwhile reviewing the work done so far. Also, the level of success achieved should be mentioned and the conditions under which the experiments have been carried out should be made clear. Further, a careful distinction must be made between particulate-gas systems, in this case dry coal cleaning and water slurry separations.

Coal-water slurries have been the subject of a considerable amount of research which has been going for several years. It can be said that for a given coal in which the inorganic matter has been liberated by grinding or perhaps other means, most of inorganic sulfur minerals can be removed by HGMS along with some part of the ash-forming mineral matter. The same cannot be said yet for dry separations.

Several attempts have been made to improve the magnetic properties of the impurities in coal so that these might be removed by liberation and subsequent magnetic separation. Siddiqui steam-treated Pakistani coal and noted that a small part of the coal could be picked up by a permanent magnet. This fraction had a higher sulfur content than the non-magnetic fraction. Yurovskii and Remesnikov used shorter treatment times on a Soviet coal and reported a sulfur reduction in a fraction beneficiated in an induced roll separator. Adamov added magnetite to coal ground to minus 3mm and achieved some ash reduction by magnetic separation.

Kester, Leonard and Wilson at W. Virginia University found that from 54 to 86% of the inorganic sulfur could be removed from untreated coal. The size range was 48 x 200 mesh and the best results were obtained on Upper Freeport and Pittsburgh seam coals. Erratic results were obtained with a 48 x 0 size fraction. The separations were performed using a Frantz Isodynamic Separator.

Ergun and Bean at the Bureau of Mines did an extensive project on pretreating and beneficiating coals. Separating untreated coals in an induced roll separator, they report reductions in the pyritic sulfur of from 7 to 60% depending on the coal. Three passes were made through the separator. A -300 micron sample of Pittsburgh coal was pretreated with a heated gas in a fluidized bed and then subjected to separation in an Isodynamic Separator. Reduction in pyritic sulfur increased from 27% before pretreatment to 32 to 45% afterwards.

This work seems to indicate that some of the pyritic sulfur can be removed from some coals provided that it can be liberated at a large enough particle size. All of these studies have used gravity feed to the separator and have been concerned at least with dry separations if not separations of material from a gas stream. Power plants blow fine coal (often -200 mesh) into their furnaces. If the pyritic sulfur is liberated, it might be interesting to see whether magnetic separation could be used on a relatively high velocity gas stream carrying coal to reduce the sulfur content.

Studies at MIT last year on an Upper Freeport coal blown through a high gradient separator showed that this kind of separation is more difficult to perform than a corresponding coal-water slurry separation. In fact, some of the best wet separation results we know of were obtained with this coal before the particulate-gas separations were attempted. Total sulfur reductions reached 57% compared to about 40% obtained by Trindade on Brazilian coal at MIT in 1973 and up to 45% by Murray on Illinois and Indiana coals recently.

It is important to distinguish between total sulfur reduction and inorganic sulfur reduction since sulfur is found in both forms in coal. The organic sulfur is generally not removed by physical means such as magnetic separation. Murray's test result in dry separation of 40% inorganic sulfur reduction would correspond to a 20% total sulfur reduction if half the total sulfur occurred as inorganic, say pyrite.

Sulfur reduction by physical separation depends on the type of coal, its history, grinding, liberation, etc., so it is very difficult to compare any of these figures. But the problem of dry separation of impurities from coal is far from solved and represents a research need and opportunity.

I cannot be optimistic enough to agree with the authors overstatement that "HGMS has provided a possible commercial method, for economically separating the sulfur-bearing and ash-forming minerals from dry pulverized coal on a large scale." There are technical problems remaining in dry separation and it is not clear what kind or how much of pretreatment might be necessary nor how the HGMS process might be applied here.

This is a fertile field for research and development and the need for such should not be obscured by statements which imply that the problem has been solved and is waiting for commercial development. This is one area still in need of research and that research ought to be done by a number of people or research groups. These should include magneticians, fluid flow experts, chemists, and other experts in surface phenomena, electro-statoc effects, etc.

There is a reference here to the "commercial applications of HGMS to coal desulfurization." We are not aware of any such commercial applications and should comment that the authors' use of "pilot-scale" often refers to laboratory experiments.

The application of modeling to HGMS has been made by several people. Reference should be made to work reported at Intermag meetings of the IEEE. In the case of coal, Trindade at MIT used a model for water-coal separations in 1973 and Clarkson has extended the fluid model to an extended range of Reynolds numbers and has taken inertia and the boundary layer near matrix fibers into account.

Beyond the generalities offered here about research needs, there are a number of specific topics that could and should be mentioned.

Agglomeration of particles may be a major problem in particulate-gas systems especially when a separation is to be performed and an effort has been made to liberate one particle species from another. Obviously, in collection systems this may be desirable but the agglomeration of material needs to be understood in any case.

The effect of fines on particle agglomeration needs to be known. Methods of grinding to reduce fines may be employed and moisture levels varied to change the effect of the fines content in the particulate system. Electrostatic charges on particle surfaces have an effect on agglomeration and are certainly affected by moisture. The carrier gas, its moisture content and temperature are all considerations for separation studies.

For matrix separations the flow characteristics of gas and particles past matrix materials should be studied over a reasonable range of Reynolds numbers. A number of studies of this sort have been done for fluids based on work such as Zebel's on aerosols but they have generally considered the fluid incompressible.

Loading of matrices in a high gradient separator will have to be studied. There are now available both batch-type and continuous separators, the latter offering a possible solution to short duty cycles in the former when they occur. Since neither coal loss nor inefficient collection of impurities is desirable in coal cleaning a clean separation is the goal. Besides agglomeration mentioned before, mechanial trapping, collection efficiency, and changes in the flow resistance, all functions of loading, affect the quality of separation.

The authors point to the work of Drehmel and Gooding as an illustration of the application of HGMS to particulate-gas separation. This is a very interesting piece of work and should be continued as a vigorous program. A point not made here though is that in the case of BOF dust all of the material wants to be trapped out of the gas stream. I would call that collection rather than separation. The difference between the two is significant. Mechanical trapping aids collection and therefore enhances the magnetic effect. In particle separations on the other hand mechanical trapping can be a serious problem whether in gas streams or liquids.

It is stated by the authors that an "observation" is made in Murray's work that confirms the enhancement of high gradient magnetic separation by the use of combined matrix materials. In fact, Murray reports only the use of Frantz screens in his experiments. Many people have patented matrix designs using carpet tacks, steel wool, ribbon screens, grooved plates, etc. In our experience, however, steel wool or stainless steel wool (not necessarily the magnetic 400 series) remains an excellent matrix material. Expanded metal is nearly as effective and is used in larger pilot plant moving matrix HGMS devices. Steel shot, filings and grooved plates have the disadvantage of occupying part of the magnetized volume of the HGMS device and magnetized volume is what one pays for.

Matrices consisting of combinations of material types have been used in batch and moving matrix HGMS devices. We found this type of matrix to be very useful in treating iron ores in a program carried both at our laboratory and on location in Minnesota. It could almost go without saying that matrix design constitutes a development used and that companies engaged in machine production are not unaware of this need.

OPEN DISCUSSION

Liu: I think I should quickly mention that our research has been totally funded by the RANN Division of NSF. They have been very helpful supporting magnetic separation, as Dr. Oder is obviously aware. I agree with most of the comments made about magnetic separation. Much of the data was not produced in our paper for this workshop because approval has not been granted by the Electric Power Research Institute.

HIGH GRADIENT MAGNETIC SEPARATION: REVIEW OF SINGLE WIRE MODELS

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INTRODUCTION

High Gradient Magnetic Separation (HGMS) has become a subject of considerable interest since its introduction into commercial beneficiation processes and the hope that it may be a means of removing certain pollutants encountered in industrial processes and energy generation. Most systems employing HGMS which are in use so far have been wet systems but a discussion of such systems may be useful in suggesting possibilities, as well as difficulties, with the use of air or other gases instead of water. Here our main concern will be with the collection process on single fibers which can be considered the basic "building block" of the more complicated systems in actual use.

In HGMS processes a collection of ferromagnetic fibers is commonly placed in a magnetic field of sufficient strength to saturate the fibers. Hence, each properly oriented fiber becomes essentially a line dipole which produces rapidly changing fields with position in its vicinity. The resultant field gradients produce a magnetic force on magnetized particles in the vicinity of the fiber. Under appropriate conditions, particles will be swept toward the surface of the fiber and will be held there as long as the field is present. The major competing forces on the particles in such systems are usually the magnetic forces which sweep the particles toward the fibers and the forces due to the fluid which carries the particles. In a properly designed system, the magnetized particles will be retained on the fibers which act essentially as a filter for these particles.

Here we shall review the various factors that are of importance in the collection process and discuss the mechanism of particle capture on single fibers. In most practical HGMS filters, presently built, water carries the particles which are to be separated out through a steel wool filter. The ability of the magnetized fibers of the steel wool to hold the magnetized particles will determine the effectiveness of the filter. Single fiber experiments to study the collection process are usually carried out with steel or nickel wires acting as collecting fibers. The reader will be referred to the appropriate literature for derivations of various quantitative models. The emphasis here will be on the presentation of detailed experimental results of particle buildup on single collecting fibers (nickel wire) for several geometrical arrangements of the fiber relative to the flow and the applied magnetic field.

FACTORS INFLUENCING PARTICLE CAPTURE

It is convenient to consider separately each of the elements entering into filter operation. Such an approach provides considerable simplification though sometimes with some sacrifice of rigor. To begin with there are the particles which are to be removed from the slurry. It is obvious that these particles must be capable of being magnetized. If the particles are ferro- or ferrimagnetic they will either possess a net magnetic moment or can be magnetized by a magnetic field. Usually the moment will be large, and, once induced by a field, may be permanent. Such particles may be conveniently removed by more conventional methods and HGMS may pose certain problems. For instance, attempts to remove the collected particles by flushing out the filter after turning off the field may not be fully successful since the particles may, in effect, become permanent magnets. Paramagnetic particles, on the other hand, have no permanent magnetic moment and a field is required to induce such a moment which is proportional to the product of the applied field and the susceptibility of the particle. It is obvious that weakly paramagnetic particles require larger fields. Diamagnetic substances have not been removed by HGMS since the very weak magnetic moment induced in them opposes the applied field so that weak, repulsive rather than attractive forces result in the usual collection region.

The magnitude of the force, considered for paramagnetic particles only in the following, is proportional to the product of the magnitude of the induced magnetic moment, the gradient field and the volume of the particle.[1,2] For larger particles the relationship becomes more complicated, partly because the field gradient can no longer be considered uniform over the entire particle volume.[3] But major interest usually centers on small particle sizes since it was the (successful) intent of HGMS to obtain sufficiently large gradient fields to make separation of micron and submicron size particles possible. At those small sizes, the major competing force is the fluid drag which tries to carry the particles along. Obviously, below some critical size the fluid force will always be larger than the magnetic force and no collection will take place. There are no particular difficulties in calculating the magnetic forces although different approaches are possible and are still the subject of inquiry. Gravitational forces and particle inertial effects may play an increasingly important role in the case of larger particles. At the other end of the scale, very small particles could be subject to thermal motion as well as other effects. Electrical forces play a role in the capture process by influencing particle to particle, as well as fiber to particle, interactions. Finally, magnetic interactions between particles can become a factor for particles with large moments.

The fiber serves both as the collection center and as a means of producing the high gradient field. It has to have appropriate mechanical and chemical properties (i.e., it should not be corroded by the water usually used as the fluid that carries the particles). Secondly, it should be ferromagnetic with a large magnetic moment so that high gradients can be achieved. The shape of the fiber cross-section is the other factor that determines the magnitude of the gradient. Hence, small diameter wires or steel fibers with relatively sharp edges are used to obtain large gradients. Steel with appropriate magnetic and corrosion-resistant properties is probably the best fiber material for many HGMS applications, but nickel can be used and is sometimes more readily available in a variety of sizes, as needed for experimental purposes.

The particle carrier, for instance the water used to make the slurry in conventional HGMS systems in use today, plays a vital role in the process since it carries the particle past the collecting fibers. A determination of the flow fields and how they affect the collection process is one of the major problems facing investigators.

Generally, the fluid flow problem has been considered under various, simplifying assumptions. At very small Reynolds' numbers, a viscous flow, also called creeping flow[4,5], has been assumed by several investigators and has been solved by a method due to Lamb. This solution has been claimed to give a better approximation of actual flow conditions close to the cylindrical fiber than other often used solutions. At somewhat larger Reynolds' numbers the assumption of potential flow is made, but is probably reasonable only some distance from the fiber. Boundary layer effects near the fiber surface have been postulated[6] to take into account the hydrodynamic forces on particles near the fiber surface. When considering the particle capture process, the calculation of forces on particles building up in layers at the fiber surface has been based on various assumptions[6,7,8,9]. In some cases, predictions of particle buildup seem to follow observed patterns, in others they are off by several orders of magnitude. The actual flow conditions such as a stagnant region and the formation of eddies and an unsteady flow region, all downstream behind the cylindrical fiber seem to make calculations more difficult, especially at larger Reynolds' numbers which occur at higher flow velocities. Experimental results will be presented to point out such difficulties. One of the references quoted above deals with the capture of particles by a dielectric fiber[4] due to electric rather than magnetic forces. When the particles are electrically neutral but possess an electric dipole moment induced by an applied electric field, then the capture mechanism becomes analogous to that of magnetic dipoles captured on a ferromagnetic fiber. Since such analogies may be useful--capture by electric fields is receiving a great deal of attention by workers studying the behavior of aerosols--a further discussion of this analogy is in order.

Generally, the induced electric dipole moment can be assumed to depend linearly on the applied field. The same assumption can be made for magnetic dipoles over a limited range of applied magnetic fields. A paramagnetic particle will have a dipole moment proportional to applied field for most practically obtainable values of field. However, magnetic fibers will saturate with sufficiently large fields whereas dielectric fibers will only show saturation phenomena if they are made of a ferroelectric substance.

Electric fields are necessarily limited in magnitude to below electric breakdown gradients, but there is no such limitation on magnetic fields. (However, the region in which corona discharges take place may be used in electric separation processes.) Hence, magnetic gradients are limited only by geometrical considerations and by the magnetic properties of the collecting fibers.

EXPERIMENTAL RESULTS

Collection of paramagnetic particles on a nickel wire has been studied by photographic means and through the use of a hot wire anemometer[10,11,12]. Here photographic collection data will be presented for the three idealized cases that are usually discussed in the literature.

Case A has the collecting fiber, flow, and the field all mutually perpendicular.

Case B has field and flow parallel to each other and perpendicular to the fiber axis.

Finally, the third arrangment, case C, has the flow along the axis of the fiber with the field again perpendicular, as it must be if the fiber-field combination is to produce a large magnetic field gradient. All three cases are shown in Figure 1.

The details of the experimental method are given in references 9-11. All experiments described here were carried out with a nickel wire as the collecting fiber ($4\pi M_s$ = 6900 Gauss for nickel) and paramagnetic particles of manganese pyrophosphate being collected from the slurry. The manganese pyrophosphate, $Mn_2P_2O_7 \cdot (3H_2O)$ is reported to have a magnetic susceptibility of 186×10^{-6} emu/oested/gram. Particle sizes ranged from about 1 μm to 100 μm with most of the particles in the 1-10 μm range, but an appreciable fraction of the total particle volume is made up of large particles.

Photographs for case A and case B have been obtained by taking pictures of the collecting wire "end on." Case C shows the wire with the attractive regions on the sides and one repulsive region facing the camera with the flow parallel to the wire. In all pictures the flow is from top to the bottom except in Figure 10.

Figure 2 shows the buildup of particles on a 25 μm diameter nickel collecting wire, for case A geometry. Parameters are 4.5 cm/s slurry velocity and a 12.5 kOe applied field. The concentration of $Mn_2P_2O_7 \cdot (3H_2O)$ is 0.0264 g/l. The time shown below each picture is measured from the instant at which the slurry first reached the wire (A flow of deionized water was carried past the collecting wire until a diverter valve changed the flow to introduce the slurry. Flow velocities were controlled by means of an orifice downstream far from the collection region).

Figure 3 shows the buildup again for case A using a 25 μm diameter wire and 0.0264 g/l concentration of $Mn_2P_2O_7 \cdot (3H_2O)$, with a flow velocity of

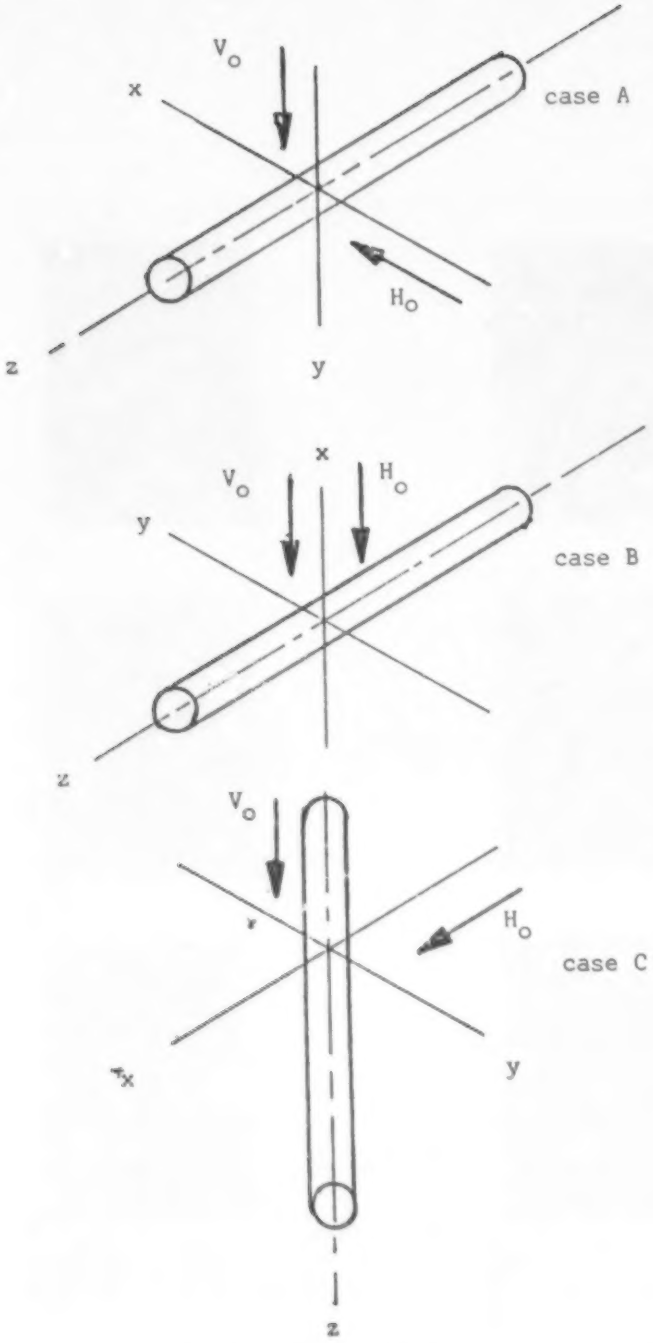

V₀
x
case A
H₀
z
y
x
V₀
H₀
y
case B
z
V₀
H₀
case C
x
y
z

Figure 1. The three cases of flow and magnetic field orientation relative to the collection wire.

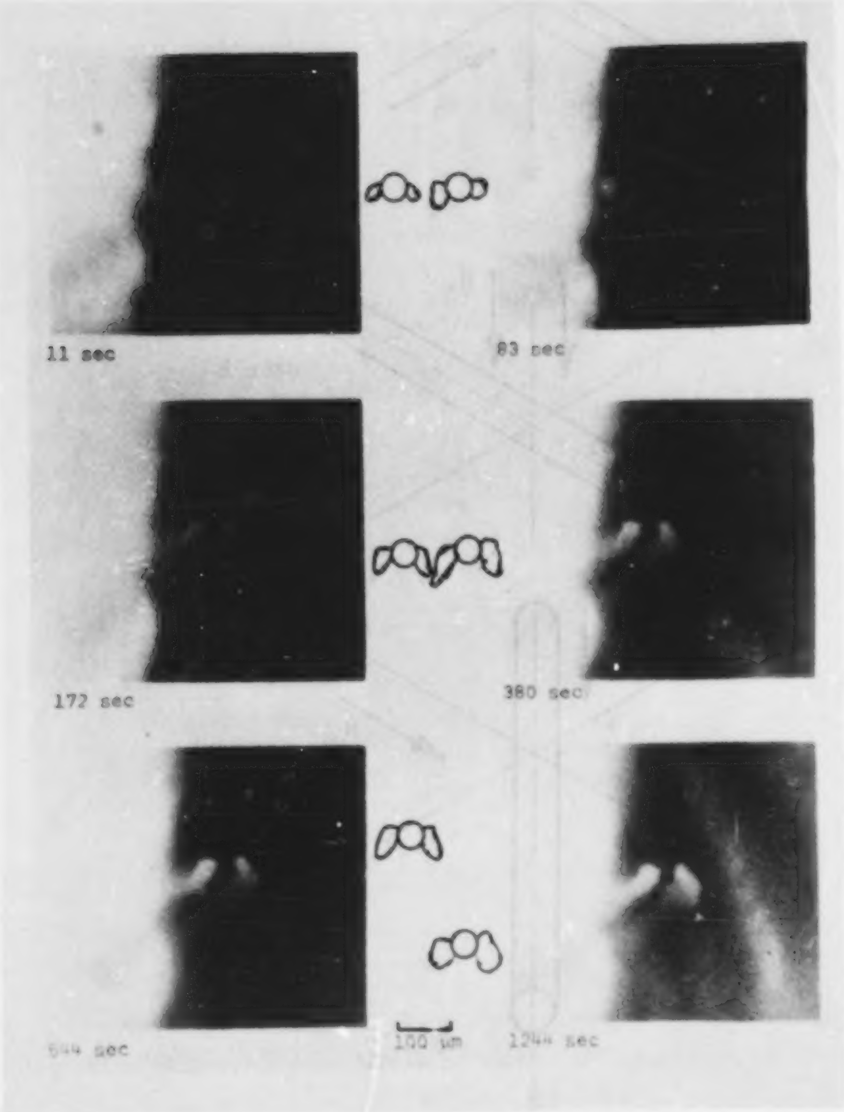

11 sec
83 sec
172 sec
380 sec
544 sec
100 µm
1244 sec

Figure 2. Buildup for case A--.0264 g/l $Mn_2P_2O_7 \cdot (3H_2O)$, field 12.5 kOe, flow velocity 4.5 cm/s, 25 µm diameter nickel collecting wire.

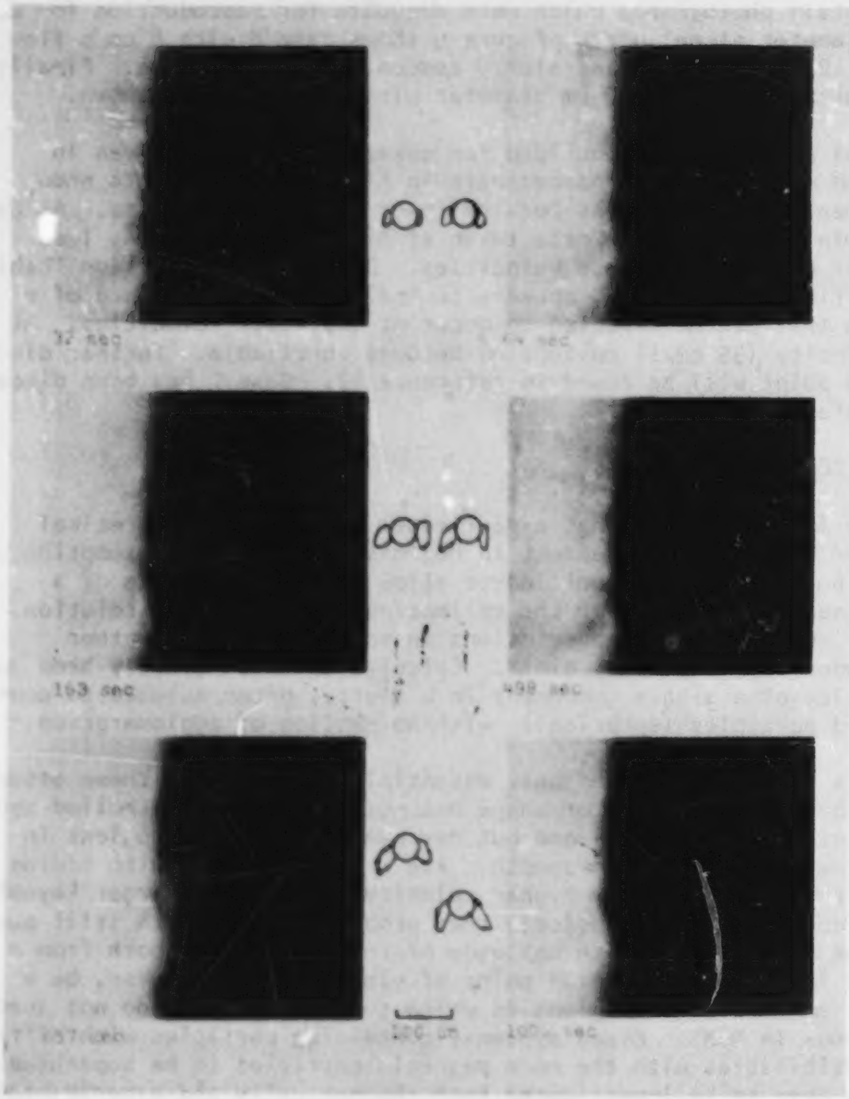

32 sec
44 sec
163 sec
498 sec
120 μm
1003 sec

Figure 3. Case A--0.264 g/l $Mn_2P_2O_7\cdot(3H_2O)$, Field 7kOe, flow velocity 10 cm/s, 25 μm diameter collecting wire.

10 cm/s and a field of 7 kOe. Figures 4, 5, 6 show the buildup on a 125 μm diameter nickel wire for case B for increasing velocities 3.7 cm/s in Figure 4, 17.8 cm/s in Figure 5 and 35.0 cm/s in Figure 6, all for a 12.5 kOe applied field. In Figures 7 and 8 a 17.8 cm/s flow velocity was used, with a 3.0 kOe field in Figure 7 and a 1.0 kOe field in Figure 8. It was more difficult to obtain photographs which were adequate for reproduction for case B with 25 μm diameter nickel wire. Figure 9 shows case B with 8 cm/s flow velocity and a 12.5 kOe field, and slurry concentration as before. Finally, in Figure 10 collection on a 250 μm diameter wire for case C is shown.

Theoretical predictions of buildup for cases A and B were given in references 6 and 11 and are reproduced here in Figures 11 and 12 to show the good agreement with experiment for the lower values of velocity. As can be seen by examining some of the data taken at higher velocities[12], less reliable data is obtained at these velocities. Downstream collection "behind" the collecting fiber, for case B, appears to indicate the appearance of a stagnant region that can be expected to occur at the lower velocities. At the highest velocity (35 cm/s) collection becomes unreliable. Further discussion on this point will be found in reference 12. Case C has been discussed in detail in references 13 and 14.

DISCUSSION AND CONCLUSION

In summary it can be said that experimental results and theoretical predictions give substantial agreement in regions in which the assumption of potential flow holds, possibly modified to allow for the existence of a boundary layer near the surface of the collecting fiber. Lamb's solution (creeping flow) may give better predictions in some cases, but further studies are needed to settle this claim. Calculations have usually been made for the separation of a single component in a slurry, often assumed to consist of uniform-sized particles (spherical), with no mention of agglomeration.

Experiments have been carried out, essentially duplicating these assumptions, except that particle size or shape has not often been controlled and, again, agglomeration is not mentioned but may play a role. In regions in which the flow may be other than "smooth," i.e., unsteady flow with eddies "behind" the wire[15], or generally higher velocity (and, thus, larger Reynolds' numbers) flow, collection data indicate that problems arise which still await a solution. The difficulties with unsteady or turbulant flow, both from a theoretical and from an experimental point of view may not, however, be a matter of major concern since regions in which such flow occurs do not seem to be advantageous in HGMS. Mixed systems, containing particles with different magnetic susceptibilities with the more magnetic particles to be separated from the slurry need to be investigated both theoretically and experimentally in single fiber systems. Questions of particle to fiber and particle to particle adhesion and how they differ in liquid and gaseous carrier systems need to be answered. Considerable differences between the two systems (wet and dry) can be expected and have apparently been observed in the operation of entire systems[16]. Instabilities may occur in the particle motion[17] and experimental observation of particle trajectories is needed to test the theoretical

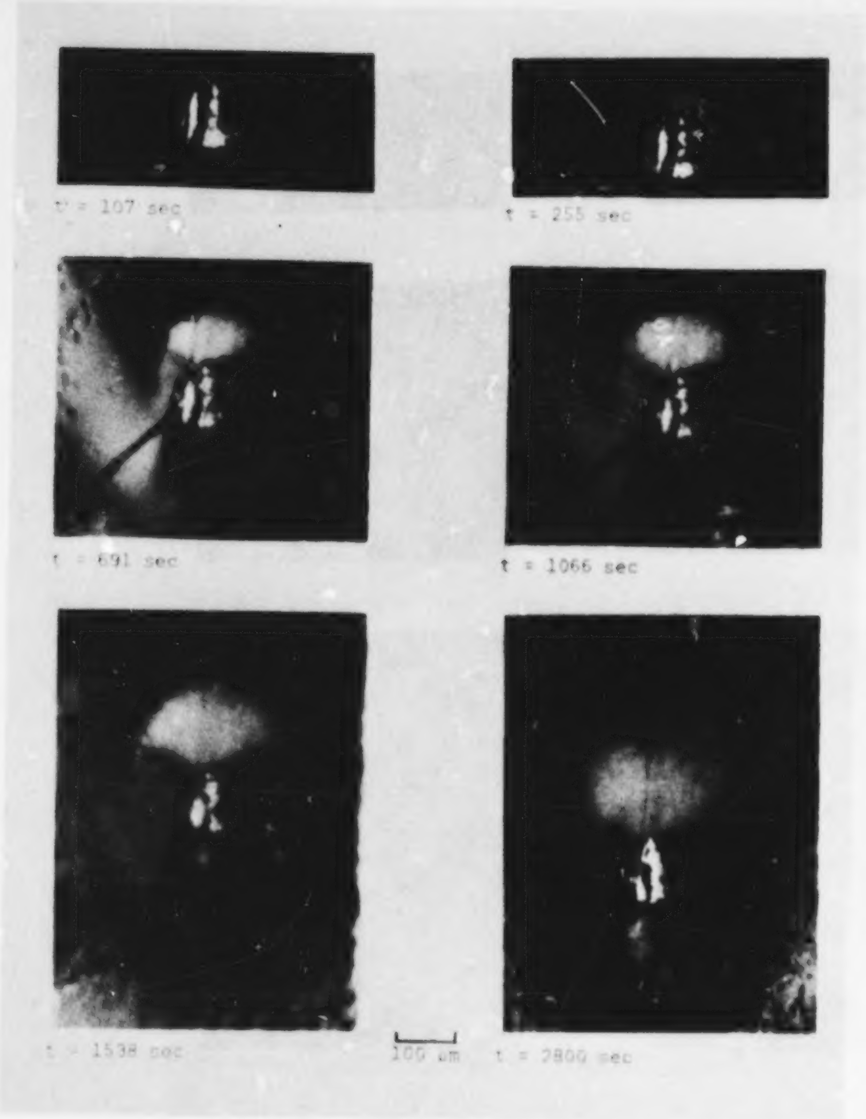

t = 107 sec
t = 255 sec
t = 691 sec
t = 1066 sec
t = 1538 sec
100 µm
t = 2800 sec

Figure 4. Case B--0.264 g/l $Mn_2P_2O_7 \cdot (3H_2O)$, flow velocity 2.7 cm/s, field 12.5 kOe, 125 µm diameter collecting wire.

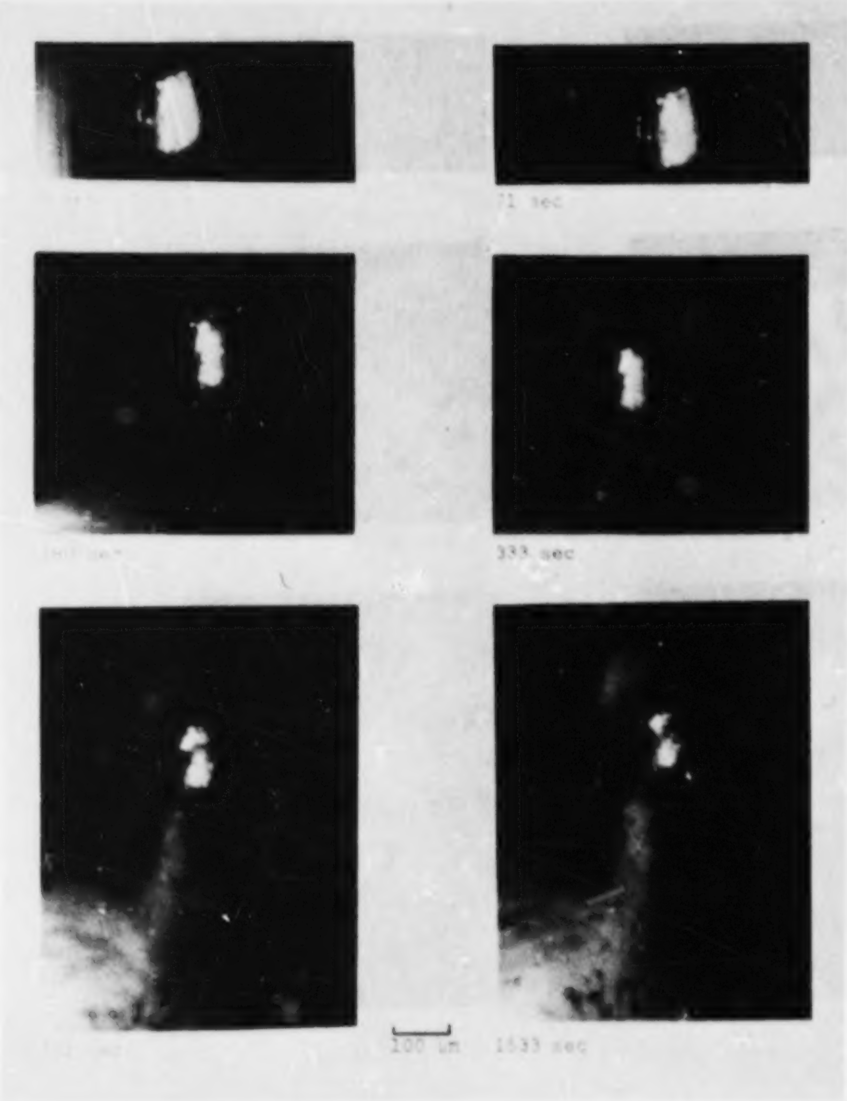

71 sec
333 sec
100 μm
1533 sec

Figure 5. Case B--.0264 g/l $Mn_2P_2O_7\cdot(3H_2O)$, field 12.5 kOe, flow velocity 17.8 cm/s, 125 μm diameter collecting wire.

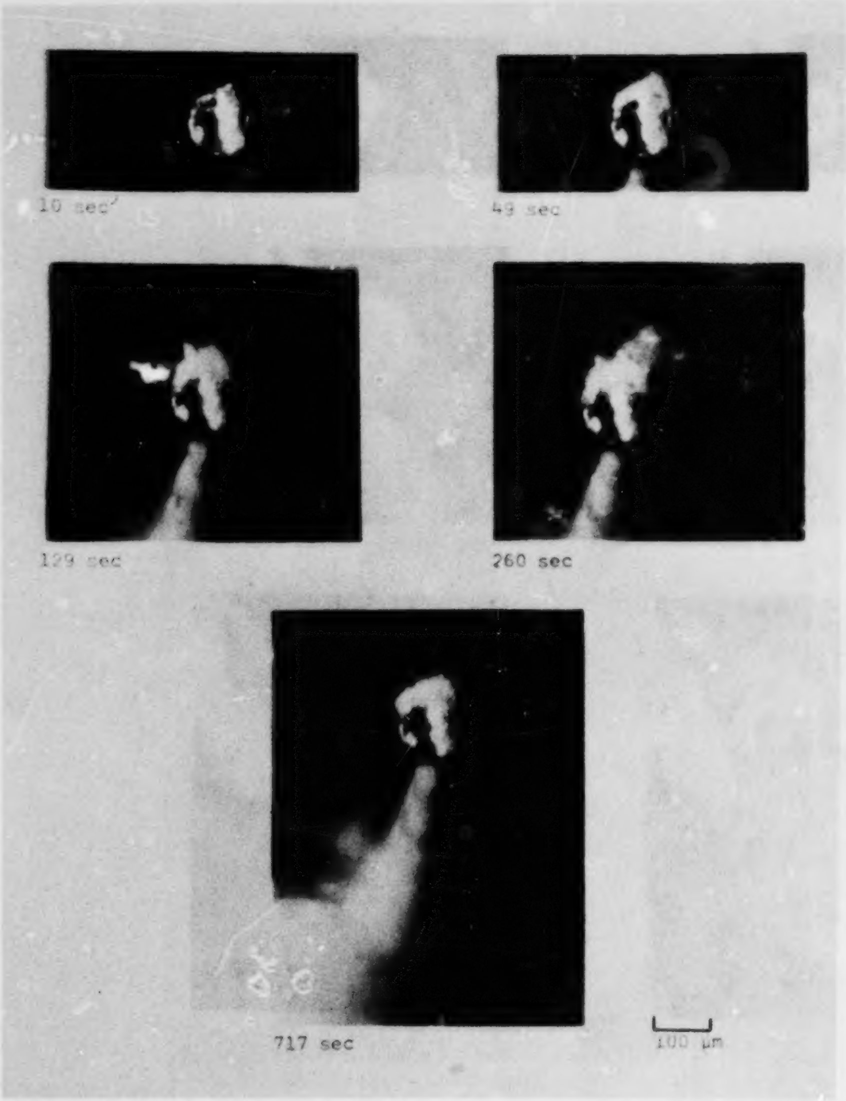
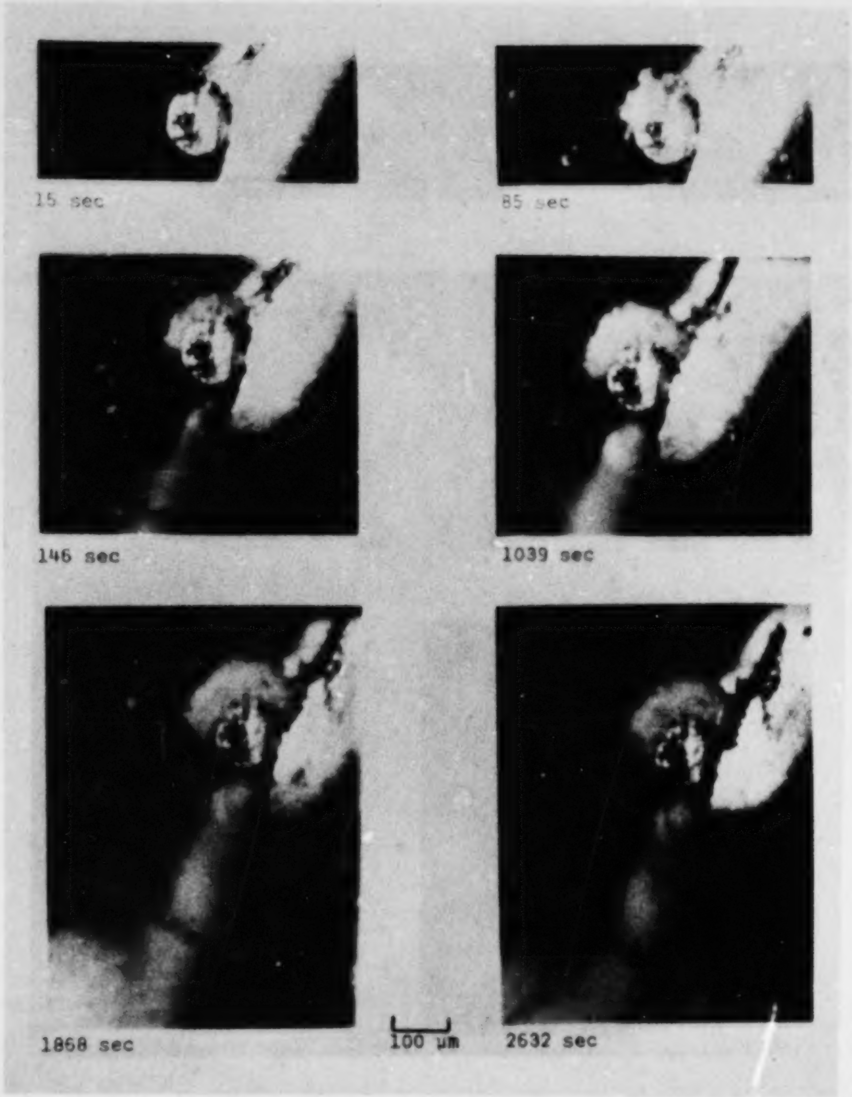

10 sec
49 sec
129 sec
260 sec
717 sec
100 μm

Figure 6. Case B--.0264 g/l $Mn_2P_2O_7 \cdot (3H_2O)$, field 12.5 kOe, flow velocity 35.0 cm/s, 125 μm diameter collecting wire.



15 sec

85 sec

146 sec

1039 sec

1868 sec

100 μm

2632 sec

Figure 7. Case B--.0264 g/l $Mn_2P_2O_7\cdot(3H_2O)$, field 3 kOe, flow velocity 17.8 cm/s, 125 μm diameter collecting wire.

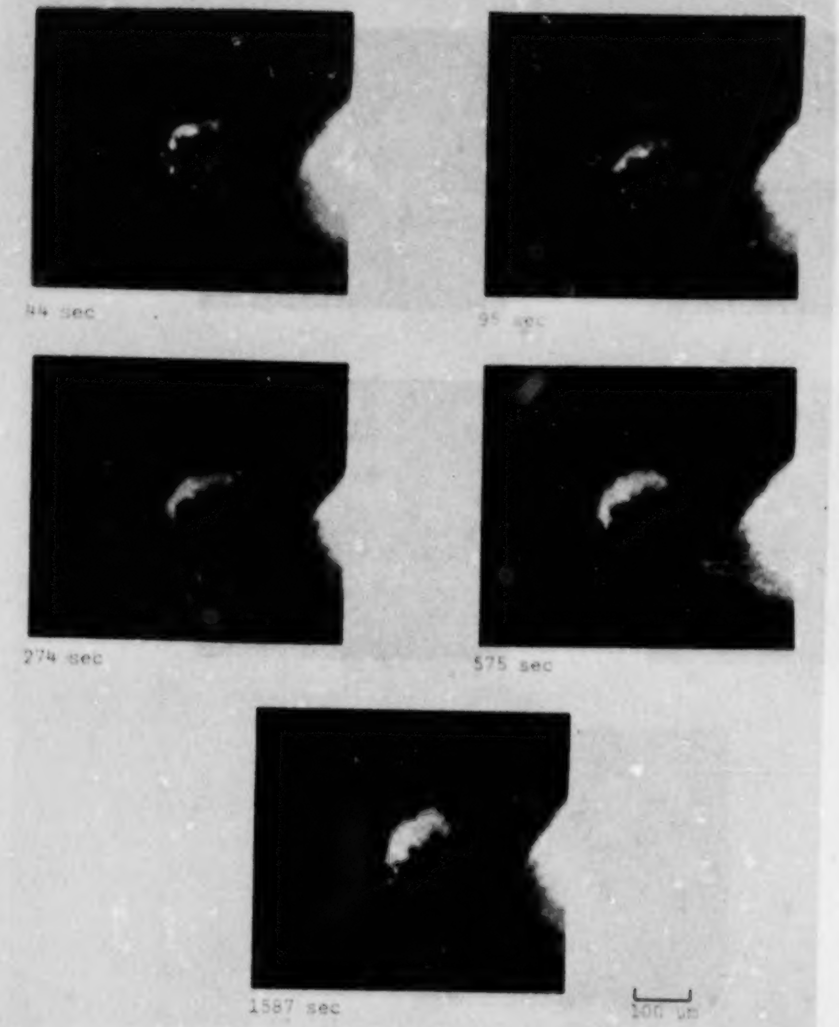

44 sec
95 sec
274 sec
575 sec
1587 sec
100 µm

Figure 8. Case B--.0264 g/l $Mn_2P_2O_7 \cdot (3H_2O)$, field 1 kOe, flow velocity 17.8 cm/s, 125 µm diameter collecting wire.

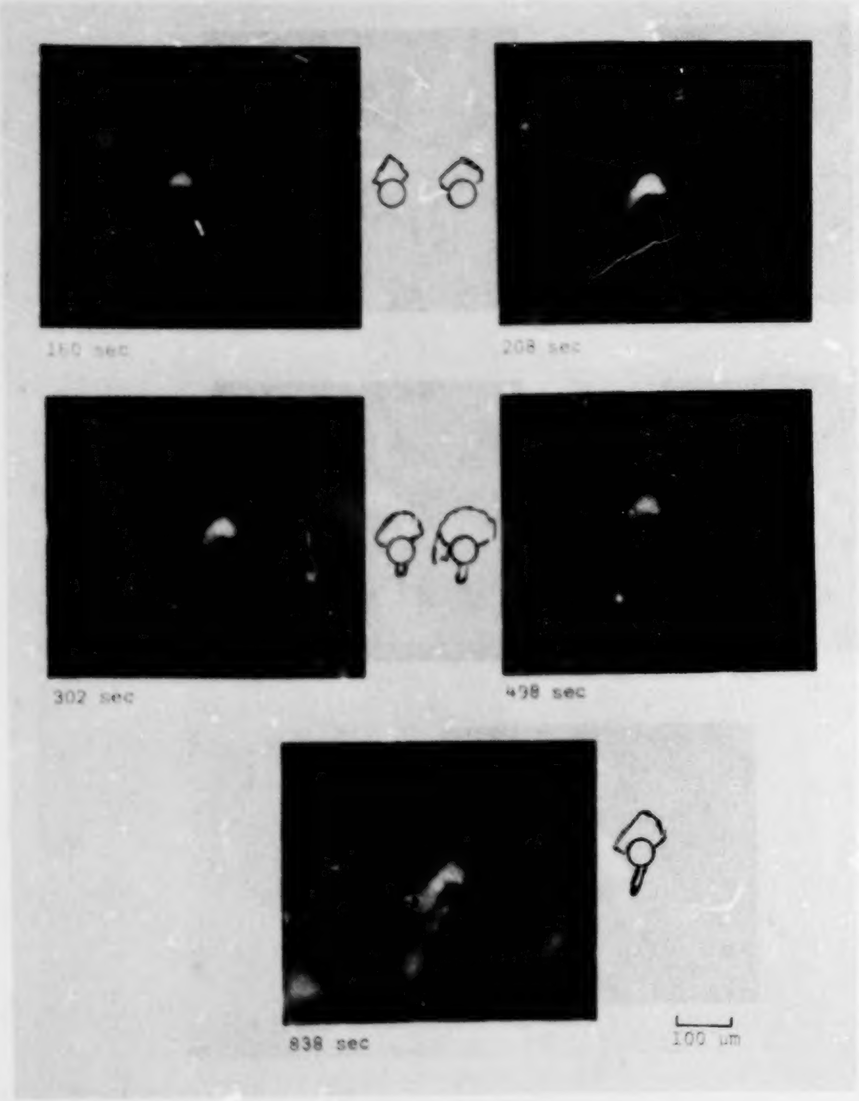

160 sec
208 sec
302 sec
498 sec
838 sec
100 μm

Figure 9. Case B--.0264 g/l $Mn_2P_2O_7\cdot(3H_2O)$, field 12.5 kOe, flow velocity 8 cm/s, 25 μm diameter collecting wire.

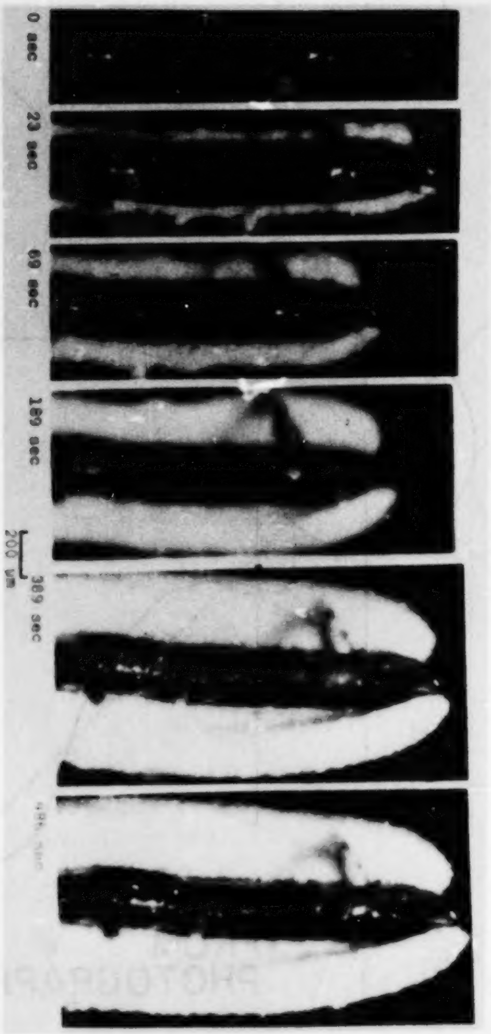
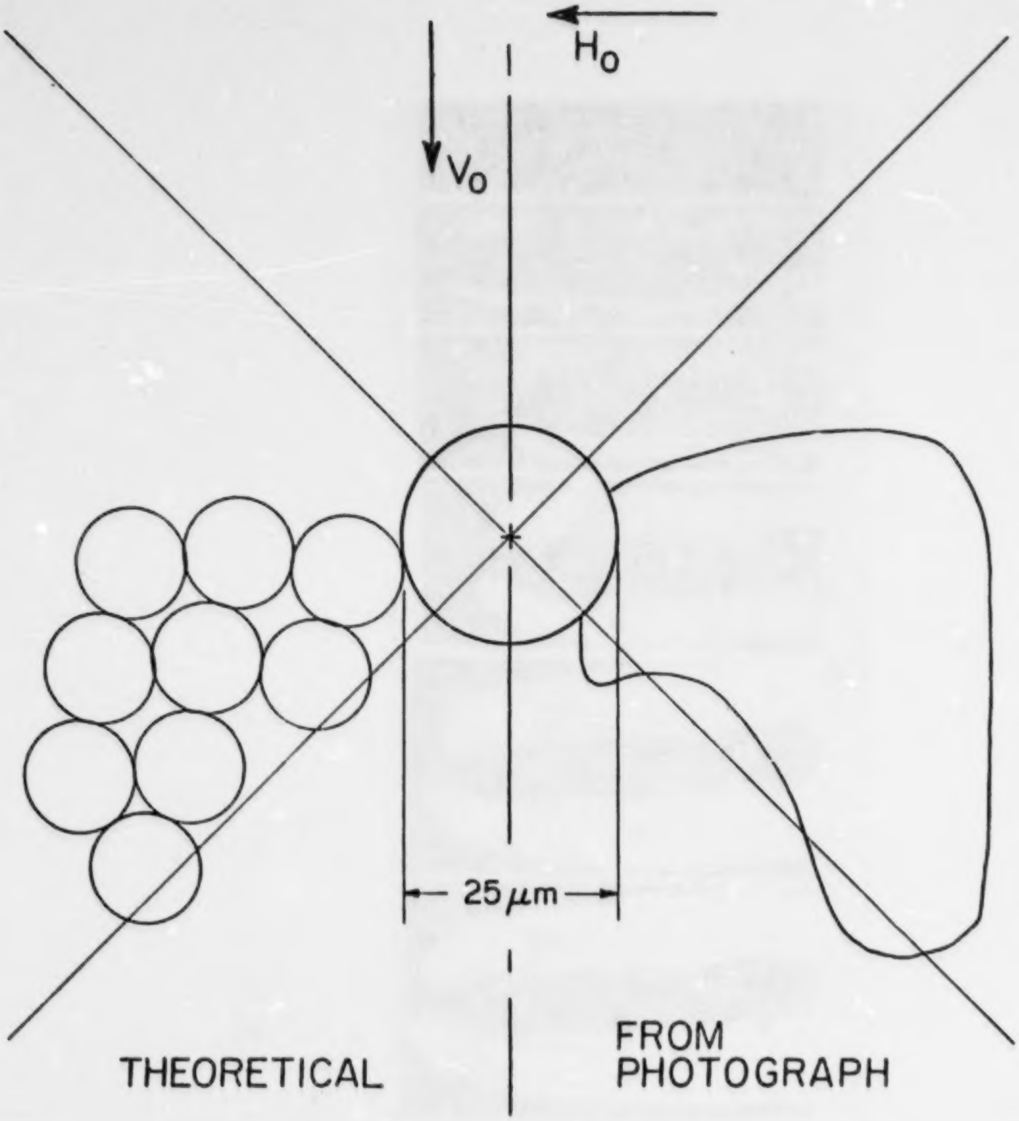

0 sec
23 sec
69 sec
189 sec
200 μm
389 sec

Figure 10. Case C--.0664 g/l $Mn_2P_2O_7\cdot(3H_2O)$, field 12.5 kOe, flow velocity 10 cm/s, 250 μm diameter collecting wire.



H_o
V_o
$25\,\mu m$
THEORETICAL
FROM PHOTOGRAPH

Figure 11. The theoretical buildup for case A compared to buildup shown in photographs.

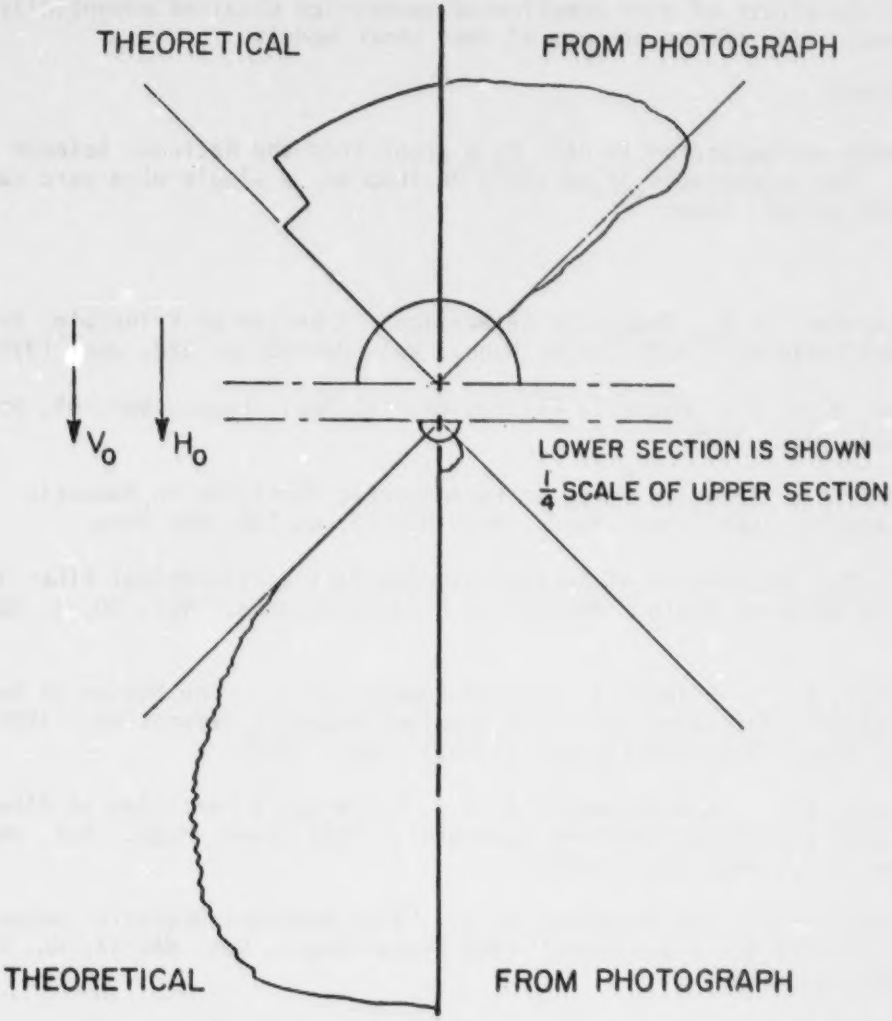

THEORETICAL
FROM PHOTOGRAPH
V_o
H_o
LOWER SECTION IS SHOWN
$\frac{1}{4}$ SCALE OF UPPER SECTION
THEORETICAL
FROM PHOTOGRAPH

Figure 12. The theoretical buildup from Luborsky[6] compared to buildup as shown on photographs for case B.

predictions. Since many systems in use today have collecting fibers arranged in other than the idealized geometries of cases A, B, or C some thought has to be given to the effect of more complicated geometries obtained essentially by a linear combination of two or more of the ideal models.

ACKNOWLEDGMENTS

This work was supported in part by a grant from the National Science Foundation. The photographs of particle buildup on a single wire were taken from a thesis by Carl Cowen.[18]

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WRITTEN DISCUSSION

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As Professor Friedlaender points out, most HGMS experiments have been concerned with the separation of particles using a liquid carrier media. In viscous fluids, particle inertia and the Earth's gravitational force can generally be disregarded when describing the particle dynamics. This is the low Stokes number regime where Watson's theory[1] is applicable. The agreement between gross filter efficiencies calculated using Watson's theory and results of actual filter experiments is generally unsatisfactory. Luborsky states in his review[2] that from the mass of material captured in randomly packed magnetic filters it was clear that Watson's theory would have to be extended to account for particulate accumulation on individual fibers. Thus, the need arose to examine the particulate buildup on single wires. Subsequently, theories were developed based on trajectory models and concurrently the photography of particulate buildup on single wires in an aqueous medium was begun at Professor Friedlaender's lab. The agreement between these theories and experimental observations has not been entirely satisfactory and, as noted in the article under discussion, only the slower flows substantiated the theoretical predictions.

The problem with the theories may be the result of the complexity of the physical situation. Professor Friedlaender correctly indicates many of the difficulties, in particular those associated with the description of the fluid flow. Different flow fields around the cylindrical fiber have been used in single fiber theories and have yielded quite different quantitative results. In addition it appears that there are conditions of practical interest for which the relative velocity between the particle and the fluid medium become large. For such non-Stokesian particles one should of course use the empirical formula for the fluid drag in the equation of particle motion.

It is perhaps worth emphasizing again that the above theories are based on a particle trajectory model. Although Himmelblau[3] has observed some HGMS trajectories, neither these observations nor other experiments provide satisfactory particle dynamical data to check this basic model. As Kolm[4] has advocated and Professor Friedlaender suggests, careful observations of the particle dynamics are needed to check the trajectory models and to provide a better understanding of the basic processes involved in particle capture.

What of particulate-gas separations using HGMS? Professor Friedlaender has noted some of the potentially different aspects of dry separation as compared to wet HGMS. Drehmel and Gooding[5] have demonstrated the efficiency of dry magnetic separation in cleaning simulated iron furnace emissions. However, in many instances there was not even qualitative agreement with fiber efficiency predictions based on Watson's theory. This is not surprising since Stokes numbers for their experiment ranged from about 2.5 to 250---consistently above the threshold marking the onset of inertial effects. A general treatment of the effects of particle inertia and the Earth's gravitational force when vectorially combined with the magnetic and fluid forces is given in Reference 6, but these results are still experimentally untested. Calculations of single fiber capture distances and particle trajectories and more details on the qualitative differences between the inertial and noninertial cases of particle dynamics can be found there. The two factors which make inertial effects manifest in dry HGMS are the reduced viscosity of the medium and the typically higher flow velocities used in gas filtration. The reduced fluid mass density can cause the Earth's gravitational pull to be more significant than with denser liquids.

Moreover, it is not yet clear that a satisfactory filter efficiency can be derived from single fiber trajectory calculations. In fact, some preliminary work on multiple fiber filter theory indicates that the effect of neighboring fibers in large arrays can be quite significant. Shown in Figure 1 are two pairs of capture efficiency curves for a long fiber presented as a

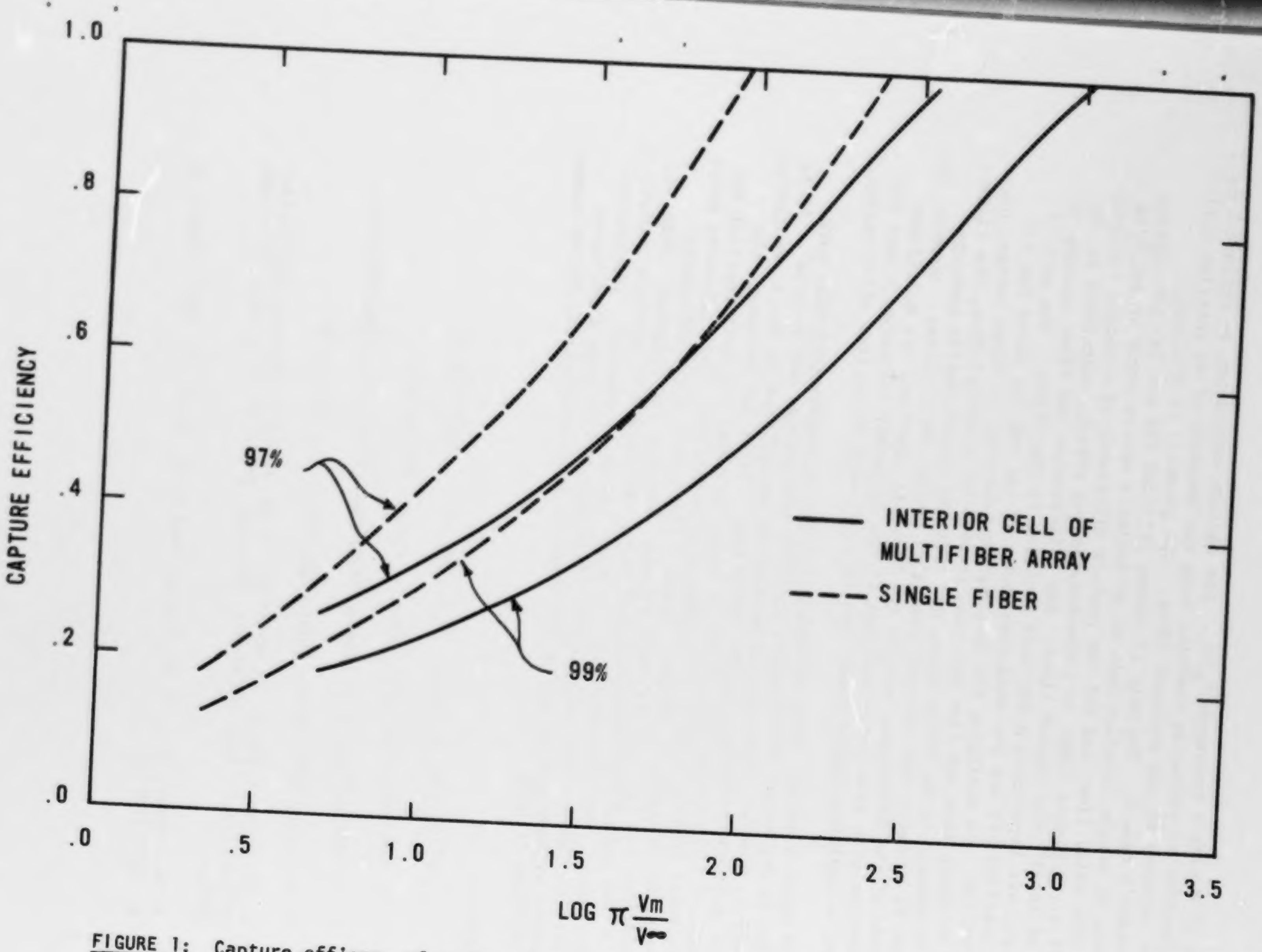

FIGURE 1: Capture effiency of an interior cell of a multifiber filter based on 97% and 99% void filter fractions using flow and magnetic field descriptions of a square multifiber array (solid lines) and an isolated single fiber (dashed lines).[7]

function of Watson's parameter V_m/V_∞. The capture efficiency is defined here as the fraction of particles starting from the boundary of an interior cell which is captured by the enclosed fiber. The geometry is Professor Friedlaender's case B. The pair of curves labelled 99% are (a) the capture efficiency of single interior cell of an infinite square array of parallel fibers (the solid line) and (b) the capture efficiency of a single isolated fiber (i.e., a single fiber in a non-interacting array), calculated by the method of Zebel[8] (the dashed line). The efficiencies and fiber spacing in the arrays are based upon a 99% void fraction in the filter. The pair of curves labelled 97% are the analogous cases of (a) and (b) above for 97% void fraction. In analyzing the information contained in these curves, one should be aware that the flow has been treated as potential flow, the viscous force is assumed to be Stokes drag, and that the square matrix geometry analyzed is probably not the most practical design for a filter. However, the magnitude of the difference between the isolated fiber and the multifiber efficiencies with even loosely-packed arrays certainly indicates a need for caution in applying single fiber results to predict multifiber efficiencies.

We agree with the lists of research needs in fundamental HGMS work compiled by Professor Friedlaender in these proceedings and by Kolm[4]. It is our belief, however, that priorities must be established in the investigation of high gradient magnetic capture mechanisms. First and foremost, the dynamics of the single fiber particle trajectory model in both inertial and non-inertial regimes should be checked experimentally. The trajectory model has been the cornerstone of most recent buildup and filter efficiency calculations and its validity has not been directly investigated. Next, in view of new multifiber data presented here, the nature of particle capture in geometrically - arrayed multifiber filters should be studied. Information provided in these areas could enable designers to make more intelligent decisions balancing engineering and economic considerations in future commercializations of HGMS.

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OPEN DISCUSSION

Ariman: If you go back through all the HGMS presentations, I have a couple of things to say. First of all, the workshop presentations and discussions so far have been very exciting. In the morning sessions we did have very interesting comments from specialists in different fields. For example, in the morning, acoustics was the main area of presentations. We had stimulating discussions from researchers in the acoustics,filtration and HGMS areas. But when we cover the HGMS area it was quite interesting to see that, all of the comments during the discussion presentation were made by HGMS workers.

I would like to make the following comments for the interest of the participants who are experts in the HGMS. It seems to me there is a strong emphasis for the basic reference in this field to Zebel's work. I think it is an excellent work. Quite a few of you have been doing a lot of work using his model. It is a single fiber, single particle model. Professor Friedlander's statement that Zebel did not only consider dielectric particles and conducting fibers, but he looked also at both conducting and nonconducting particles. For one of the cases, he obtains a closed form solution for the particle trajectory equation. But for the other case he has a numerical solution for the particle trajectory.

He also looks at both viscous and non-viscous flow cases. He has published a total of four articles on this subject. Two were coauthored by one of his associates, Walkenhorst. They also did some experimental work. But they talk of a very special type of fiber filter. In other words, this fiber had a special formation and a lot of distance between the fibers. That single fiber model will hopefully be suitable for that particular filter. However, I really do not know how good that single wire model would be for the very complicated filter which is used in the HGMS process.

On the other hand we have modified and extended the single fiber model to include the interference effects between the fibers. In the model there are an infinite number of rows and infinite number of fibers on each row. We found out that when the distance between the fibers is somewhat small, there was quite a bit of difference between Zebel's model and our model. Furthermore, a relatively recent publication, Ionia and his co-authors did both analytical and experimental work and their experimental results compare very well with our analytical model findings. Furthermore they concluded that when porosity of the filter increases, there is a decrease in the collection efficiency. Those are the results we obtained . We found the same result through our analytical model prior to their publication.

THE INFLUENCE OF ELECTROSTATIC FORCES AND OF THE PROBABILITY OF ADHESION FOR PARTICLE COLLECTION IN FIBROUS FILTERS

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INTRODUCTION

Fibre filters have long been used industrially to separate solid or liquid particles from gases. The number of publications on this topic is correspondingly large. The answer to some important questions are nevertheless obscure to the present day. This is particularly true for the technically interesting particle size range above 0.5 - 1 µm and for Reynolds' numbers larger than 0.1. Exceptionally striking differences between theoretical calculations and experiments have been observed in these regions.

Generally speaking the distances between the fibres in these filters are large compared with the particles. The particles are therefore not deposited by a screening action but must be moved to the fibre surface and retained there. It has already been shown |1| that the deposition efficiency is the product of the collision efficiency η times the probability of adhesion h.

$$\varphi = \eta \cdot h \qquad (1)$$

One important reason for the above mentioned discrepancy is the fact that the significance of the probability of adhesion has been rather neglected, i.e. it was assumed that all particles striking a fibre would also stick there. However, this is frequently not so. Many theoretical papers on transport processes, on the other hand, do not take sufficient account of physical realities and simplify the problem unduly.

In order to elucidate some of the open questions on transport and adhesion processes we have carried out a three-part research programme: theoretical and experimental studies of single fibres, model filters and industrial filters. Some of the results will be discussed below.

THEORETICAL MODEL

The deposition efficiency is calculated on the basis of a model cylinder on which the incident flow is at right angles.

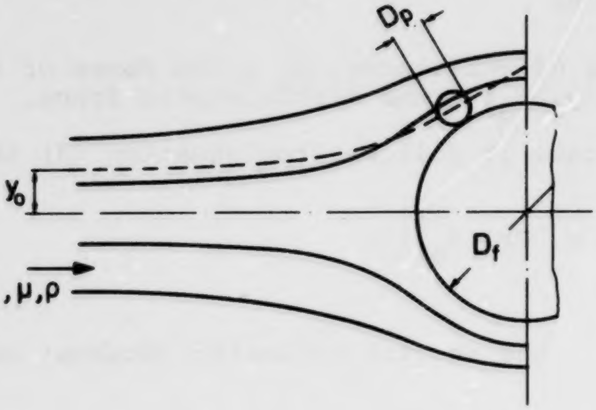


Fig.(1) - Model deposition on a fibre

The Brownian motion can be neglected in the particle size range $D_p > 0.5$ μm under consideration here. In this case the other possible transport mechanisms include: inertia, gravity and electrostatic forces. Let the collision efficiency in these cases be defined as (cf. Fig. 1):

$$\eta = \frac{2y_o}{D_f} \tag{2}$$

where y_o is the origin of the coordinate of the limiting particle trajectory in the undisturbed flow in front of the cylinder, i.e. the trajectory of particles which still just touch the fibre. All particles of size D_p which arrive in the band $2y_o$ will strike the fibre.

In order to calculate the collision efficiency it is necessary to find the limiting particle trajectory by solving the equation of motion of the particles. The general formulation is: the sum of all the forces shall equal zero:

$$\vec{F}_d + \vec{F}_i + \vec{F}_g + \vec{F}_{el} = 0 \qquad (3)$$

where F_d = the force of resistance, F_i = the force of inertia, F_g = force of gravity, F_{el} = the electrostatic force.

In the general case it follows from equation (3) that

$$\eta_{TES} = f(\psi, \mathrm{Re}, W, \mathrm{Fr}, N_{el}) \qquad (4)$$

where $\psi = \frac{\rho_p D_p^2 \cdot U_o}{18\ \mu\ D_f}$ the inertia parameter (Stokes' number),

$\mathrm{Re} = \frac{U_o \cdot D_f \cdot \rho}{\mu}$ the Reynolds number, $W = \frac{\rho_p}{\rho}$ the density ratio,

$\mathrm{Fr} = \frac{U_o^2}{g \cdot D_f}$ Froude's number and N_{el} an electric charge parameter.

N_{el} is the ratio of the electrostatic force at a certain point to the drag force. Various modifications of the electrostatic effect are in principle possible, depending on whether the particles, the fibres, or both of them are charged and whether external fields are applied. Assuming that both the particles and the fibres are charged, i.e. that Coulomb forces act,

$$N_{el} = N_{Qq} = \frac{Q \cdot q}{3\pi^2 \cdot \varepsilon_o \cdot \mu \cdot D_f \cdot D_p \cdot U_o} \qquad (5)$$

where Q is the fibre charge per unit length, q = particle charge, ε_o = influence constant = $8.86 \cdot 10^{-12}$ Coul/V·m.

The differential equation of the particle trajectory can only be solved numerically. The required condition of contact is that the particle surface must touch the surface of the fibre.

Most authors simplify the calculation according to equation (3) by entering only a single transport effect and estimating the overall effect by superimposing various partial ones. Particles are also frequently treated as mass points, and their geometrical size is additively superimposed as what is sometimes called the interception effect. The following results will show that both these simplifications lead to errors.

The solution of the equation of motion involves a considerable problem: the velocity field for the flow round the fibre is required. There are no closed solutions for the Navier-Stokes equations for the region of Reynolds' numbers 0.1 < Re < 100 which is relevant for many practical, industrial cases. Suneja and Lee |2| worked out a numerical solution for discrete Re numbers (1, 10, 50). All that is otherwise available are approximate solutions, e.g. Lambs' solution for the viscous flow round a single cylinder, or that of Kuwabara for a fibre system who neglected the forces of inertia completely in the flow; this solution therefore holds strictly only for Re = 0.

For Re ≥ 100 the equations for potential flow are often used for single cylinders. Any decision on how closely these various approximations reproduce the true flow round single fibres, even layers of fibres, must be arrived at by careful experiments.

SOME NUMERICAL RESULTS FOR THEORETICAL SINGLE FIBRE COLLISION EFFICIENCIES

A. No electrostatic forces ($N_{el} = 0$)

We calculated particle trajectories for Re ≥ 50 with the potential flow field plus an approximate equation for the boundary layer flow |3|. For Re < 1 we used Lambs flow field.

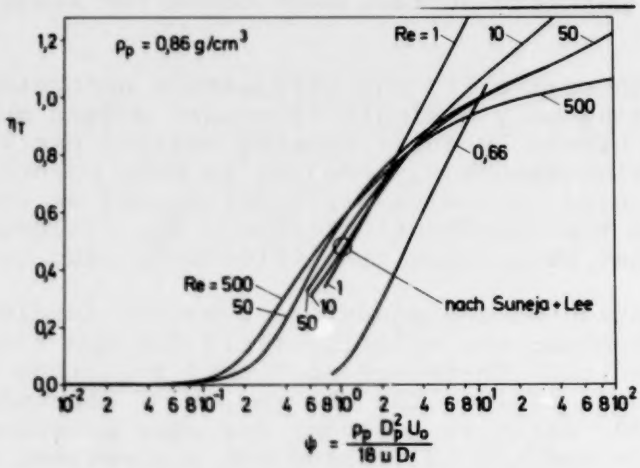


Fig. (2) - Theoretical single fibre collision efficiencies η_T (Inertia effect only)

Fig. 2 shows collision efficiencies η_T calculated for an inertia effect only i.e. (without gravity) for a particle density $\rho_p = 0,86\ g/cm^3$. In the high Reynolds number-range only the curves for Re = 50 and 500 are plotted. The curves for Re = 1, 10, 50 represent results of Suneja and Lee |2|. We see that their and our results for Re = 50 are in fairly good agreement.

For Re = 500 we found good corrrspondence with the results of Langmuir and Blodgett in the range of $\psi > 2$. Langmuir and Blodgett used only the potential flow field and their results are not plotted in the figure. For $\psi < 2$ our values are somewhat smaller, which is due to the influence of the boundary layer flow.

It is possible to fit our numerical results for $50 \le Re \le 500$ with good accuracy by the following equations:

$$\eta_T = \eta_1\,(\psi,\ Re) + \eta_2\,(\psi,\ Re,\ \rho/\rho_p) \qquad (6)$$

$$\eta_1 = \frac{\psi^3}{\psi^3 + f_1(Re)\psi^2 + f_2(Re)\,\psi + f_3(Re)}$$

$$\eta_2 = D_p/D_f = 3\,\left|\frac{2}{Re}\cdot\psi\cdot\frac{\rho}{\rho_p}\right|^{1/2}$$

$$f_1(Re) = -\,0.0133\ \ln Re + 0.931$$
$$f_2(Re) = +\,0.0353\ \ln Re - 0.36$$
$$f_3(Re) = -\,0.0537\ \ln Re + 0.398$$

The curves of Suneja and Lee show that for $\psi < 1$ η_T decreases with decreasing Reynolds number, whilst for $\psi > 1$ we observed the reverse tendency. This is because to keep ψ constant we need a bigger particle size D_p with decreasing Re i.e. with decreasing velocity U_o. This results in a higher interception effect; therefore η_T increases.

The same effect we found also in our calculations in the range $Re < 1$ using Lambs flow field.

Fig. 3 shows the results for the Reynolds-numbers 0.1, 0.2 and 0.66. There are two groups of curves. The dotted lines give the results which have been calculated by assuming the particles to be mass points ("without interception effect") whilst the full lines give the results when the geometrical size of the particles was allowed for.

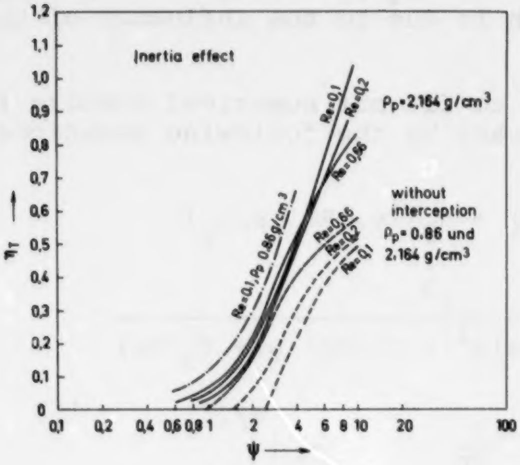


Fig. (3) - Theoretical single fibre collision efficiencies η_T, (Inertia effect, without gravity, Lamb's flow field)

From this example it will be seen that the neglect of the particle dimensions leads to considerable errors, particularly in the range $\psi > 1$. The correct contact condition must be used from the very start.

The influence of gravity was calculated for a vertical flow downwards, i.e. for a flow in the same direction as gravity is acting.

Fig. 4 shows curves plotted for different Reynolds-numbers and for two particle densities $\rho_p = 2.164$ g/cm^3 and o.86 g/cm^3. The dotted lines have been calculated as before and the particle dimension is neglected.

We see that for the correct contact condition we have high values for the collision efficiencies η_{TS}. The increasing η_{TS} with decreasing Reynolds number or decreasing particle density is again due to the fact that for constant ψ we have bigger particle sizes D_p for decreasing Re or ρ_p.

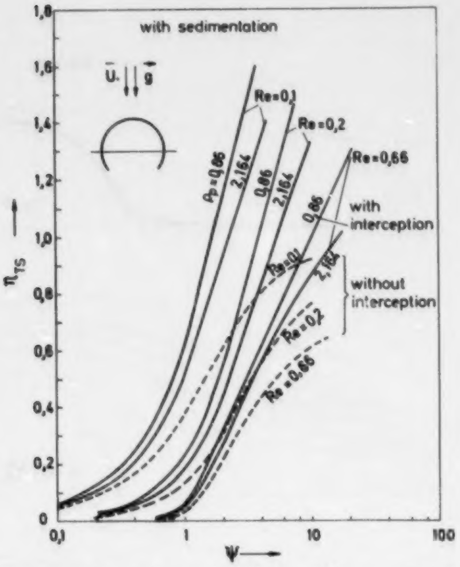


Fig. (4) - Theoretical single fibre collision efficiencies η_{TS} (Inertia effect , with gravity , Lamb's flow field)

In our opinion this effect of gravity has not been considered sufficiently till now.

B. With electrostatic forces

We must expect considerable effects in the case of coulomb forces. For this case we calculated the influence of electrostatic forces by inserting them into the equation of motion and solving the complete equation numerically. Most important are the results for the range Re < 1 which will be discussed in the following.

Fig. 5 shows some calculated particle trajectories round a cylindrical fibre for different particle sizes in the diameter range 1 - 10 µm.

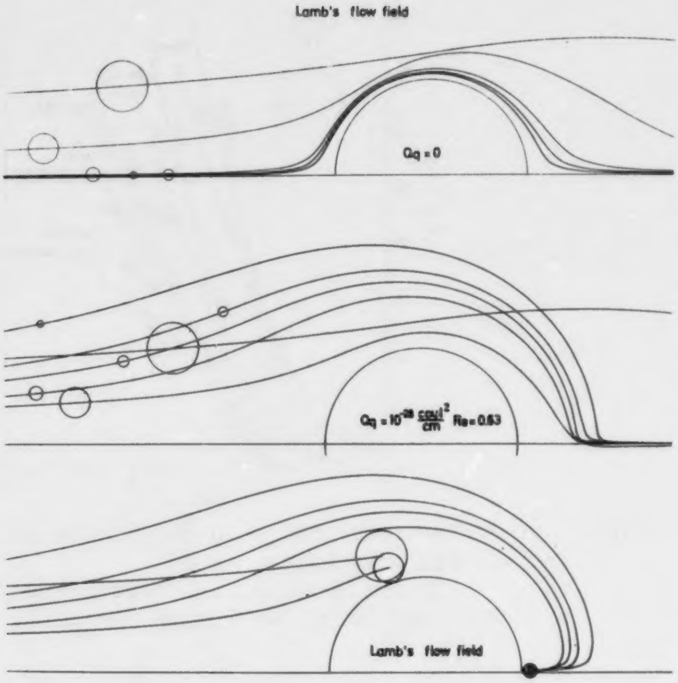


Fig. (5) - Theoretical particle trajectories
(Lamb's flow field. Re = 0.63)

The upper two examples show trajectories which just pass the fibre. All particles of the same size starting at a point 1/1000 D_F closer to the axis of symmetry will touch the fibre surface. This touching trajectory is called the limiting trajectory.

The top picture shows the case without electrostatic forces. Here collision is only due to inertia. Small particles pass round the fibre, i.e. they have small collision efficiencies. The 10 µm-particles are only little influenced by curved flow around the fibre.

In the centre we see the corresponding particle trajectories for charged particles and fibres. These have a product of the charges $Q \cdot q = 10^{-28}$ Coul^2/cm. This value is approximately equal to our experimental values.

The enormous influence of the electrostatic forces can be seen clearly especially for the smaller particles. Their collision efficiencies are increased. Obviously there exists a minimum collision efficiency at a particle size of about 5 µm. The path of the 10 µm-particle is nearly unaffected by the electrostatic forces.

The focussing of the particles on the axis of symmetry behind the fibre is a most interesting effect. This nonhomogenous distribution of particles behind the fibre has some influence of course on the deposition in the following layers of fibres. This is in contrast with any theory which assumes homogenous particle distribution.

In the lower picture we see the touching and thus limiting trajectories. The larger particles which exhibit little influence by the electrostatic forces strike the fiber on its front side. The smaller ones first pass the fibre and are then attracted to its far side.

From the limiting particle trajectories we calculate the theoretical collision efficiencies, shown in fig. 6.

In fig. 6 the single fibre collision efficiency η_{TES} has been plotted, as a complete solution of equation (3), against the inertia parameter ψ for a particle density $\rho_p = 2.164$ g/cm^3 and for Re = 0.2. The parameter of the curves is the charge parameter N_{Qq}. The above, correct contact condition was assumed for the curves marked "with interception effect", i.e. the geometrical size of the particles was allowed for, while they were regarded as mass points in the curves "without interception effect".

The curve for $N_{Qq} = 0$ holds when there is no charge. As N_{Qq} rises, the collision efficiency increases quickly for small ψ values. However, the curves for N_{Qq} pass through a minimum at about $\psi = 1$ and then approach the curve for $N_{Qq} = 0$ asymptotically. This means that above $\psi = 1$ the influence of the Coulomb force becomes less than that of the force of inertia. The total collision efficiency in the region of the minimum is less than that produced by electrostatic attraction alone, i.e. inertia and the force of gravity counteract the electrostatic force. If the force of inertia and the force of gravity were to be neglected, the electrostatic attraction by itself would give a collision efficiency independent of ψ.

This shape of the curves shows that partial solution for the equations of motion obtained by neglecting individual forces cannot be superimposed. That means that all the effective forces must be entered in the equation of motion, which should really be a matter of course but has generally not been done in the past.

In order to cover the additional influence of electrostatic forces as against the forces of inertia and gravity, the difference between η_{TES} and the collision efficiencies was formed for the case η_{TS} with no charge.

$$\eta_E = \eta_{TES} - \eta_{TS} \tag{7}$$

This difference can be approximated by

$$\eta_E = \frac{1.22(2 - \ln Re)}{\left(\frac{\psi}{2\sqrt{Re}}\right)^{1.5} + 1} \cdot N_{Qq} \tag{8}$$

This equation determines for the first time the dependence of the additional collision efficiency η_E on the parameter ψ, Re and N_{Qq}. Past publications always gave η_E no more than η_E = const $f(N_Q)$, which is fundamentally incorrect.

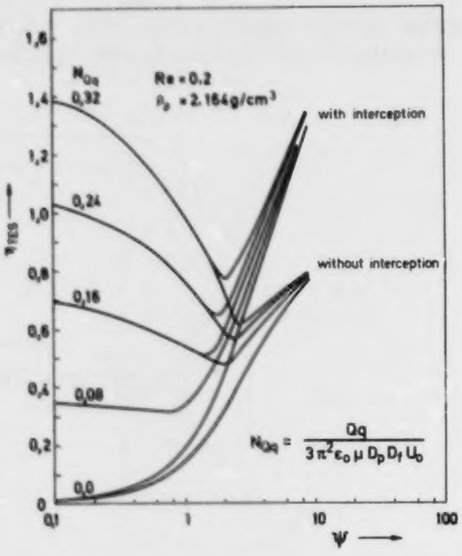


Fig. (6) - Theoretical single fibre collision efficiencies η_{TES}. (Inertia with gravity with electrostatic forces, Lamb's flow field)

DEPOSITION EFFICIENCY MEASUREMENTS

Comprehensive experiments with model filters were carried out as a check of the collision theories |5|.

Fig. 7 shows a schematic diagram of the apparatus.

The model filters consisted of several layers of parallel fibres. The fibre diameter was D_f = 38 µm. The distances between the fibres varied between 115 µm and 1600 µm in the direction of the flow and between 115 µm and 275 µm at right angles to this direction so that the percentages of the fibre volume ranged between 0.3 and 8 %.

Monodisperse NaCl particles and paraffin wax particles in the range of about 1 to 10 µm were used.

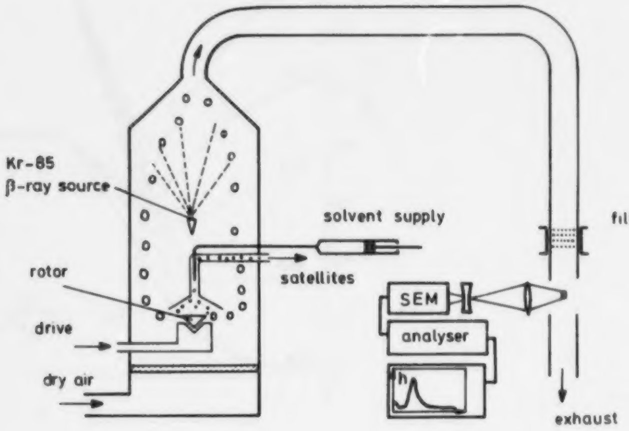


Fig. (7) - Schematic diagram of experimental apparatus

For the production of the NaCl-particles we used a spinning-top-Generator as shown in Fig. 7. The geometric standard deviation σ_g was for all size distribution < 1.12. The Kr 85-β ray source served for reducing the electrostatic charge of the particles in certain experimental runs.

Paraffin wax particles were produced by a Sinclair-La Mer-Generator. These particles are almost spheres and practically without electrostatic charge; their size distributions have geometric standard deviations $1.09 \leq \sigma_y \leq 1.12$.

Particle size distribution and concentration are measured continously and directly in the undisturbed flow by means of a scattered light instrument. This instrument developed in our institute does not require the extraction of a sample and is also capable of measuring high particle concentrations up to 10^6 Part./cm^3.

The velocity of the incident flow was varied between 25 cm/s and 200 cm/s. This results in a range from Re = 0.63 to Re = 5.2.

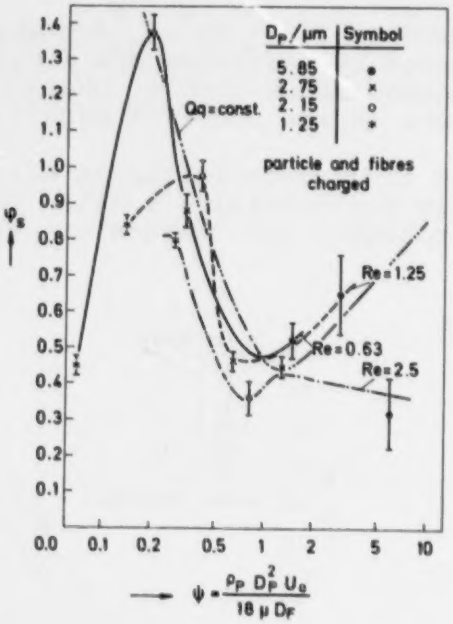


Fig. (8) - Single fibre deposition efficiencies for NaCl particles, particles and fibres charged

These measurements determine the deposition efficiency i.e. the product of the collision efficiency η and the probability of adhesion h. In Fig. 8 the experimental single fibre deposition efficiencies φ_s for NaCl-Particles have been plotted against ψ for three different Reynolds' numbers.

The shape of the curves is very reminiscent of the theoretical curves in Fig. 6. The strong maximum in the region $\psi < 1$ for particle sizes $D_p < 3$ µm is very remarkable. The deposition efficiencies measured in this region can only be explained by the effect of Coulomb forces. This is also shown by the theoretical curve for $Q \cdot q = \text{const} = 7.8 \cdot 10^9\ e_o^2/\text{cm}$ and for Re = 0.63 which has been entered for comparison as sequences of two dots and a dash.

The charge product $7.8 \cdot 10^9\ e_o^2/\text{cm}$ was confirmed by measuring the charge on particles and fibres; e.g. for 2.7 µm particles we measured about 1100 e_o charges, the fibre charge per unit length was in the order of magnitude of $10^6 - 10^7 e_o/\text{cm}$.

The fact that electrostatic forces significantly affect the curves, especially for particles smaller than 3 µm, was also demonstrated by changing the charges on particles and fibres.

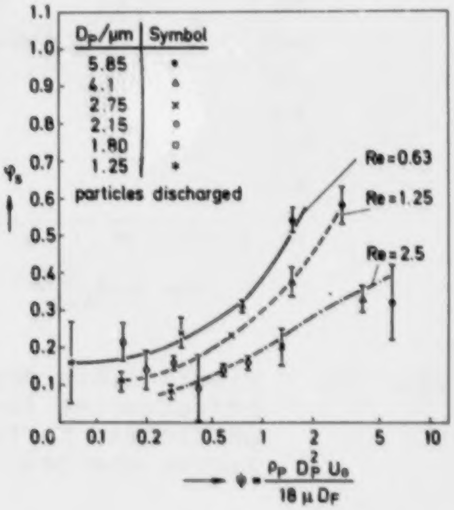


Fig. (9) - Single fibre deposition efficiencies
for NaCl particles.
(Particles and fibres discharged)

In Fig. 9 some single fibre deposition efficiencies have been plotted against the inertia parameter ψ for three different Reynolds-numbers. The corresponding velocities were in the range 25 cm/s to 100 cm/s. Compared with the results in Fig. 8 we find a drastic change in deposition especially for values of $\psi < 1$, i.e. for fine particles ≤ 3 µm. But presumably in this range particles and fibres were not completely discharged; in the range $\psi < 0.2$ the experimental values of deposition efficiencies are somewhat higher than the theoretical collision efficiencies. We suppose this to be an effect of remaining charges.

Particles of 5 µm and more were about equally deposited whether they were charged or not. This observation again confirms the theoretical prediction of a decrease in the influence of the electrostatic charge as the effect of inertia increases.

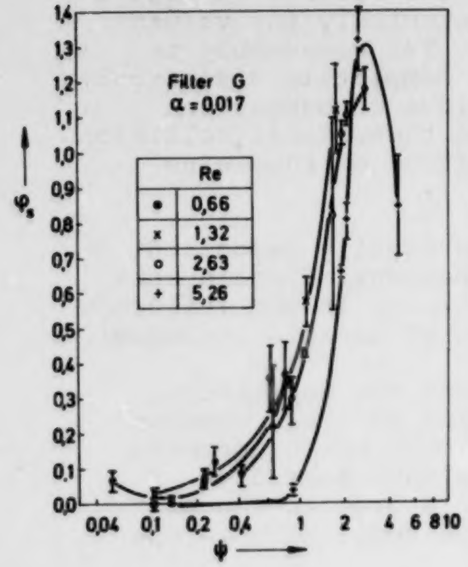
In the region $\psi > 1$ it is striking that the deposition efficiencies become smaller with increasing Re. This contradicts theoretical expectations. We explain it by an increase in the elastic rebound of the crystalline NaCl particles striking the fibre which is equivalent to a drop in the probability of adhesion. This process will be dealt with in the next section.

A completely different behaviour from that of the NaCl particles is shown by the softer paraffin wax particles. These electrostatically neutral particles were deposited much better than the NaCl particles at large inertia parameter ψ.

As we see in Fig. 10 the deposition efficiency decreased steeply with dropping ψ. The Reynolds number had practically no observable influence between Re = 0.63 to Re = 2.63. This, incidentally, is also in agreement with the theoretical prediction when the size of the particles is taken into account. A decrease in the deposition efficiency was only measured for Re = 5.26, corresponding to $U_o = 2$ m/s, and for smaller Reynold's numbers at $\psi > 2$-3, and here, too, the explanation is a decrease in the probability of adhesion.

With regard to the influence of the filter fibre volume fraction packing density α it may be noted that no dependence was found for $\alpha \leq 3$ %. The fibres behaved like single fibres.

A slight rise in the deposition efficiency was only measured at α = 5.4 %.



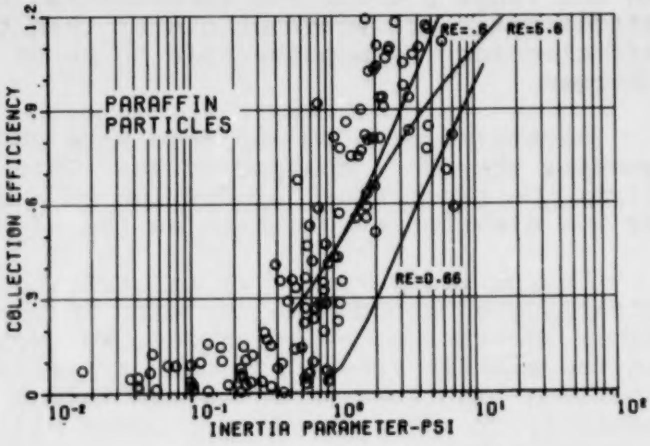


Fig.(10) - Single fibre deposition efficiencies for paraffin wax particles. Fibre Volume fraction α = 0.017

Fig.(11) - Comparison of single fibre deposition efficiencies and theories for paraffin wax particles

In Fig. 11 the experimental values of single fibre deposition efficiences for paraffin wax particles for all fibre volume fractions and all Reynolds-numbers have been plotted. The full lines represent theoretical calculations of Suneja and Lee |2| when they are recalculated for Re = 0.6 and Re = 5.6.

When a particle strikes a fibre surface it can only stick if, firstly it does not rebound and, secondly, if it is not subsequently detached.

Earlier studies |8| have shown that once particles are deposited on a fibre they stick very tenaciously. What is crucial therefore is the impact process when the particle strikes the fibre which is an elastic-plastic impact involving elastic forces of resilience which may cause the particle to rebound.

The speed at which the start of rebounding may be expected can be estimated by means of an energy balance |3,9|.

From this we obtain with the assumption that the deformation at the contact point is inealstic and that the main adhesion mechanism is due to van der Waals forces:

$$U_c = \frac{1}{k \cdot D_p} \; \frac{h^* \omega}{8\pi^2 \cdot z_o^2} \left(\frac{3}{H \cdot \rho_p} \right)^{1/2} \qquad (9)$$

Here U_c = critical velocity
k = coefficient of restitution
$h^*\omega$ = Lifshitz-van der Waals energy
h^* = $h/2\pi$ where h = Planck's constant
z_o = minimum distance of surfaces on contact ≈ $4 \cdot 10^{-8}$ cm
H = micro-hardness of the softer material
ρ_p = particle density

If we assume the deformation to be an elastic one we also obtain a $U_c \propto (k \cdot D_p)^{-1}$ porportionality. The coefficient of restitution involves the material properties but it is substantially unknown for microscopic particles. We are therefore currently working on a project for determining the impact number.

From estimates for k and other material parameters we obtained, for example, for D_p = 10 µm, U_c = 2-15 cm/s and for D_p = 5 µm, U_c = 4 - 10 cm/s.

However, these estimates do not supply the distribution of the probability of adhesion but merely the start of the rebound. More accurate information on the probability of adhesion must be obtained experimentally. We therefore developed a method for the direct determination of the probability of adhesion by the high-speed cinematograph |1,9| of particles striking a single fibre. We regard this direct method as more accurate and reliable than the comparison of measured deposition efficiencies with theoretical collision efficiencies as suggested recently by Stenhouse et al. |10|.

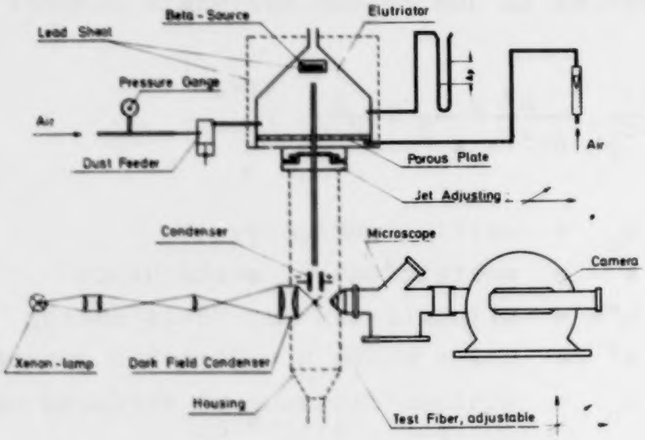

Beta - Source
Elutriator
Lead Sheet
Pressure Gange
Δp
Air
Air
Dust Feeder
Porous Plate
Jet Adjusting
Condenser
Microscope
Camera
Xenon - lamp
Dark Field Condenser
Housing
Test Fiber, adjustable

Fig. (12) - Scheme of apparatus for determining probability of adhesion

Fig. 12 shows a scheme of the apparatus by which we observed the particle trajectories.

In the apparatus a free jet, loaded with particles, impinges vertically downwards upon a fibre. The observed space is illuminated with a 250 W xenon lamp which is pulsed to three times the power during the periods of film exposure.

In order to reduce the electrostatic charge on the particle impinging on the fibre, from a β ray source, ions are generated, which accumulate on the particles. An electric field is applied at right angles to the direction of flow and this defelcts the charged particles sufficiently to stop them from colliding with the fibre.

The particle trajectories for determining the adhesion probability are recorded by the arrangement shown in Fig. 12 in which the high-speed film camera is at right angles to the axis of the fibre (max. speed 30.000 frames/second). The particle velocity can be controlled between 3 cm/s and several m/s, and particles as small as 3 μm can be recorded.

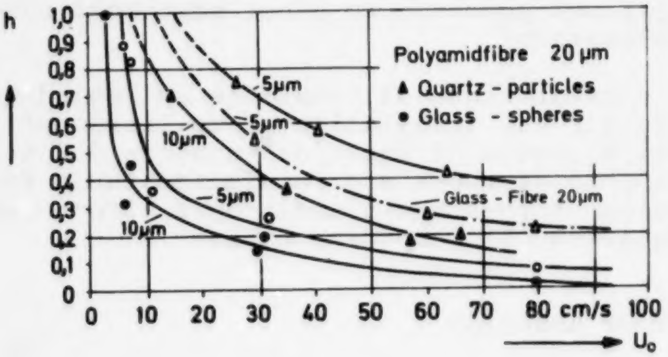


Fig. (13) - Probability of adhesion for quartz particles and glass spheres on a 20 μm polyamide fibre

Muhr's theoretical curve for Re = 0.66 was calculated with Lamb's flow field. The comparison with the readings and with the Suneja and Lee curve for Re = 0.6 suggests that at Re = 0.66 Lamb's flow field evidently provides unduly small collision efficiencies.

Below $\psi = 0.2$ the readings are higher than would be expected from the pure effect of inertia. This can, however, be explained with the Stenhouse theory |6| through elecstrostatic forces between uncharged particles and charged fibres.

Nguyen and Beeckmans |7| derived an empirical equation from measurements, but this provides much smaller deposition efficiencies than our measurements with paraffin wax particles. These authors also found a strong dependence on the Reynolds number and the percentage of the fibre volume. Presumably the discrepancy arises from the fact that Nguyen et al. based the fitting of their equation mainly on measurements with crystalline particles. In these cases the material property results in differences in the probability of adhesion - rather as with our NaCl measurements. We therefore consider the application of this empirical equation to other practical cases to be highly problematical.

In the current state of knowledge we regard Suneja and Lee's theory for the calculation of collision efficiencies in the transition region of Reynolds' numbers $0.5 < Re \leq 50$ as very suitable as long as there are no electrostatic forces. In the case of Coulomb forces their additional effect can be estimated with equation (8) according to Muhr.

PROBABILITY OF ADHESION

It has already been pointed out in the introduction that particles transported to the surface of the fibre must also be retained there. That this is by no means always the case is particularly clearly shown by the experiments with NaCl particles but also by those with paraffin wax particles. The literature contains numerous references to inadequate adhesion, e.g. the experiments of Gillespie and of Whitby are typical.

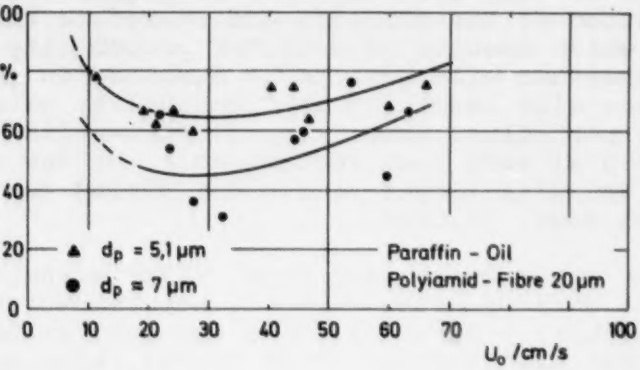


Fig. (14) - Probability of adhesion for droplets of paraffin oil

By way of example the probability of adhesion h has been plotted in Fig. 13 against the speed of incident flow U_O for various kinds and sizes of particles and types of fibres.

Rebounding starts at about 5 - 15 cm/s. As the speed increases, the probability of adhesion drops quickly. In accordance with theoretical expectations the probability of adhesion is less for D_p = 10 µm than for D_p = 5 µm, and the particles also rebound more strongly on the harder glass fibre than on the polyamide fibre. The values for the start of rebounding and the trend with respect to particle size and fibre hardness correspond fairly well with theoretical estimates.

The lower adhesion probability with glass spheres is presumably an effect of the contact geometry. Irregular quartz particles may have multi-point or area contacts while only single-point contacts are possible with the very smooth glass spheres.

The readings for the probability of adhesion for oil droplets shown in Fig. 14 demonstrate that 7 µm droplets are more prone to rebound than 5 µm droplets. It may be surprising that at 30 cm/s as many as 35 to 50% should rebound. So far we explain it by the fact that in this velocity region the droplets react as if they were elastic. At higher velocities the plastic deformation of the droplets and therefore the contact area are larger which results in a higher probability of adhesion. Still, this behaviour will have to be examined in more detail. Currently we are also examining the probability of adhesion of paraffin wax particles. Here, too, first results indicate rebounding which had already been recognizable, at any rate at higher velocities and with larger particles, during the deposition experiments on model filters.

Our own studies with model filters and real filters, just like the work of other authors, have shown that the problem of the probability of adhesion is not only important for single fibres but also for layers of fibres. With model filters |5| and NaCl particles very similar distributions of the probability of adhesion were found as for quartz particles on single fibres, even when the percentage of fibres by volume was 8%. The reason must be that in many cases the acceleration zones for particles between the fibres are often smaller than the mean fibre distances and that the particles, after a first rebound, will therefore strike subsequent fibres at approximately the same velocity. Raczynski et al |11|, for instance, have reported on studies of the deposition of quartz particles in glass sphere filters. These authors, too, found probabilities of adhesion of less than 1, even for 1 μm particles.

So far no final description of the probability of adhesion is possible because the investigations have not yet been concluded. Nevertheless existing results already explain many of the phenomena observed in the laboratory and in practice, and they also yield useful suggestions for the development or preparation of filters. Some questions are still left open but the need to allow more thoroughly for the probability of adhesion to solve practical problems is nevertheless evident.

FINAL OBSERVATION

The theoretical and experimental investigations presented here assume simple models: a single fibre and a model filter consisting of parallel fibres. In reality filter practice is quite different, at least with respect to filter structure. Nevertheless we selected this way not only because it corresponds to the procedure in the classic filter theory but also because it was possible to achieve very informative detail on the basic phenomena with the influences separated.

In a third step, of course, investigations of real filters are necessary and these we have started already. The aim of this new work is to find out the relationship between model experiments and real filters. In doing so the material properties of the particles are to be allowed for, especially with respect to adhesion properties.

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WRITTEN DISCUSSION

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The effects of electrostatic forces and the coefficient of adhesion on particle deposition in fibrous filters are among those areas which are still not well understood in filtration theory. Professor Löeffler's paper will surely contribute to further understanding of these two interesting areas of filtration and will serve to stimulate future studies.

Two parameters are important in the design of fibrous filters: the efficiency of collection and the pressure drop across the filter. Practical fibrous filters usually have high porosity, with the inter-fiber distance large as compared to the fiber diameter, and their fiber packing density is normally less than 0.1. Fibrous filters can be grouped into two classes[1]: very fine filters of high efficiency with fiber diameter on the order of 20 µm or less and moderate to low efficiency filters with larger diameter fibers of up to about 200 µm. High efficiency filters are normally used to remove submicron particles and are operated with relatively low gas velocity of a few centimeters per second. Therefore, the Reynolds number (Re) is normally less than 1. For moderate to low efficiency filters which may be used as pre-filters in ventilating systems, higher gas velocities of about 1-3 m/sec may be used and the Re could go up to about 50. For filters operated at low flow velocity, i.e., Re < 1, those deposition mechanisms due to diffusion, inertial impaction and interception are understood, and various modern filtration theories[2,3] can be used to predict the collection efficiency with good agreement with experimental data when inhomogeneity of a real filter is taken into account from the measurement of the pressure drop across the filter. Still not well understood in filtration by fibrous filters are deposition mechanisms such as electrostatic attraction, gravity effect, acoustical loading effect, the phenomenon of particle rebound at higher flow velocity, etc.

The first step in solving the basic problem of filtration is the calculation of the flow field associated with filters. Because of the complicated structures of a real filter, a filter model has to be used. Numerous such models and their calculated flow fields, such as an isolated cylinder[4], a cell model[5,6], a staggered-array model[7] and a fan model[2], have been reported. Figure 1 shows the staggered-array model used by Yeh[7] and Figures 2-3 show the flow field calculated by solving full Navier-Stokes equations numerically for two Reynolds numbers, 0.4 and 30, with fiber packing density of 0.1. The results of these flow field calculations indicate that both the Lamb's flow and potential flow describe poorly for a system of cylinders. They also suggest that Kuwabara's flow field is a good approximation to the staggered-array model for Re less than 1, even though the Kuwabara's flow was derived from Re=0. The same conclusion has been obtained experimentally by Kirsh and Fuchs[8]. It has been shown that the cell model by Kuwabara and the fan model both describe adequately the dependence of pressure drop upon fiber packing density and face velocity for Re < 1. For Re > 1, there is still no analytical solution of flow field in either a system of cylinders or an isolated cylinder.

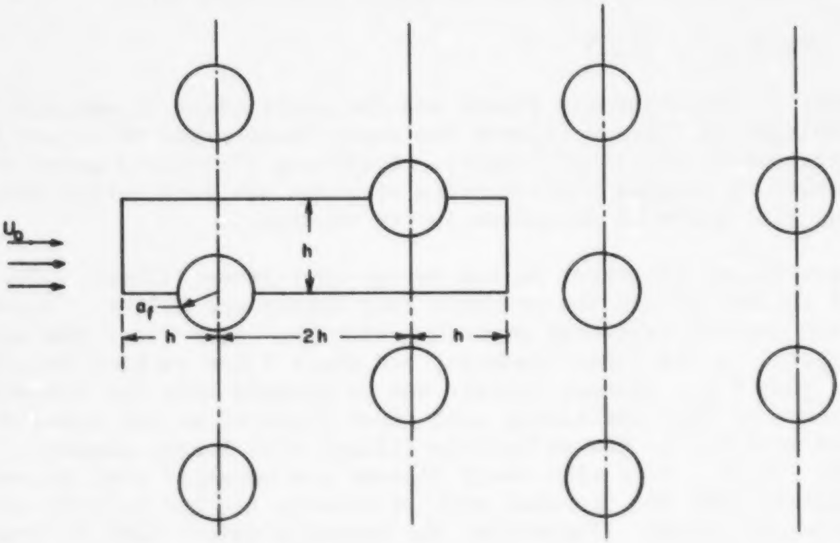

Fig. (1) Staggered-Array Model

Many current theories of aerosol filtration by fibrous filters have been based on or modified from the Kuwabara-Happel flow field for the cell model. Various theories which take into account two or more deposition mechanisms simultaneously have been developed for calculating the single fiber efficiency of a filter. These are diffusion with interception[9], inertia with interception[2,3,10] and diffusion plus inertia plus interception[3]. It would be interesting to compare results of inertia with interception presented here with other theories, but the theoretical treatment of inertia with or without interception

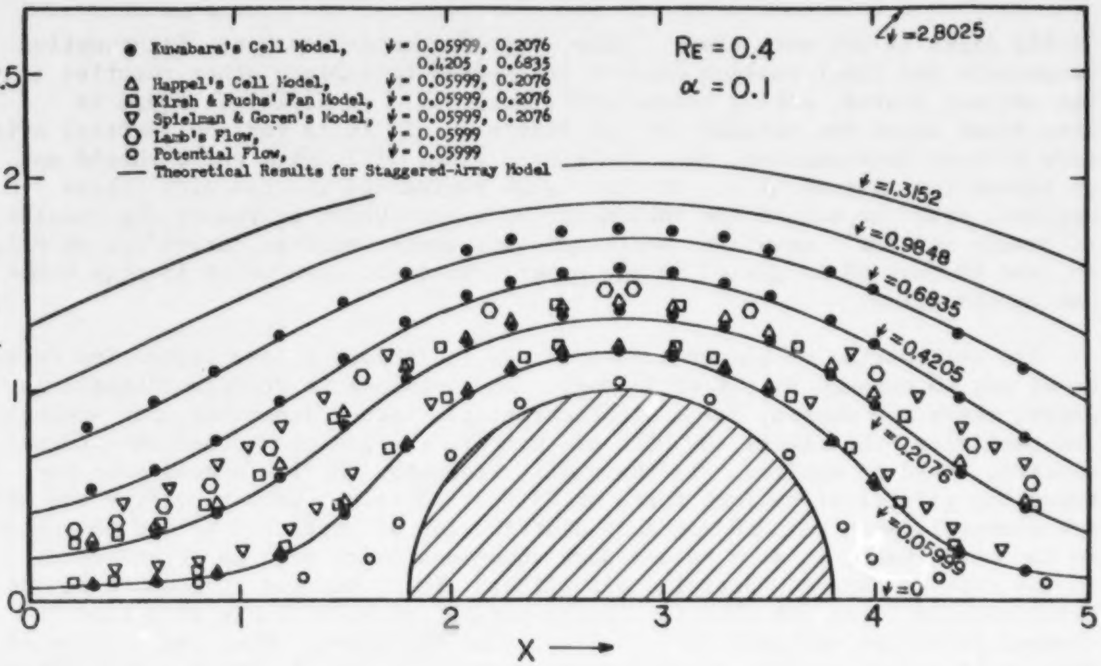

Figure 2: Comparison of Streamlines for Various Theories (Re = 0.4 and Fiber Packing Density α = 0.1)

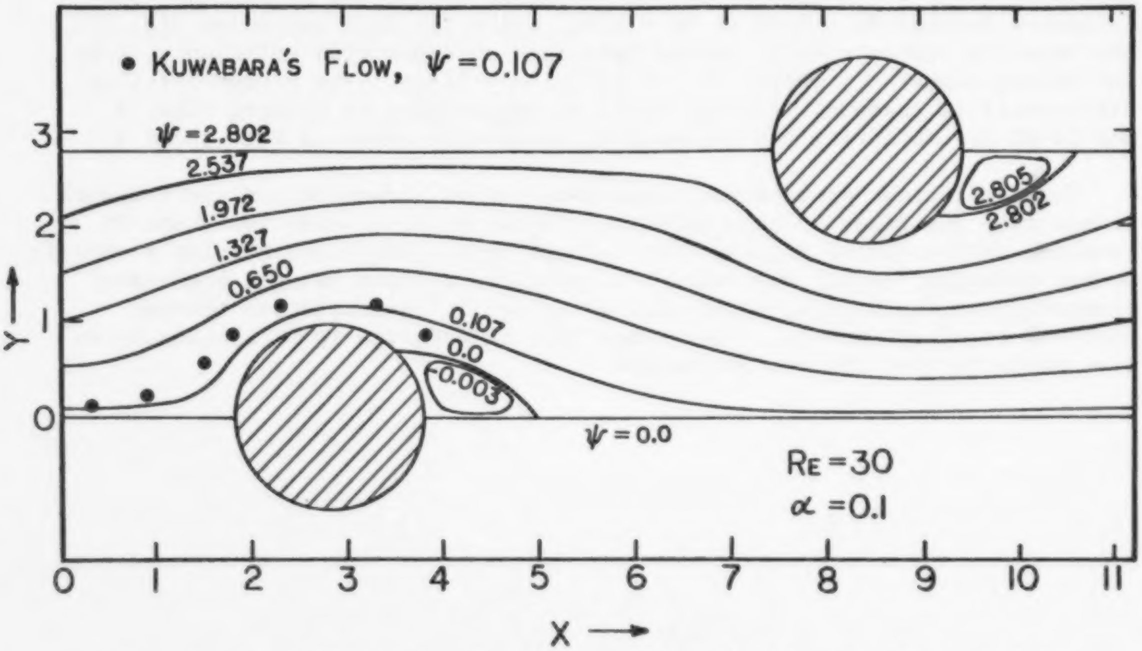

Figure 3: Streamlines for Re = 30 and Fiber Packing Density α = 0.1

in his paper is not very clear. Also, some parameters such as interception parameters and fiber packing density are needed to explore other theories but are not listed, making comparison impossible. In addition, there is some doubt about the validity of his Figure 2. If it is for the inertial effect only without interception, then the single fiber efficiency curve should not go beyond the maximum of 1. If the figure represents inertia with interception, then the method for obtaining the curves which represent the results of Suneja and Lee[11] should be mentioned, for their original curves for Re = 1, 10, and 60 (not 50 as quoted in the paper) represent results of inertia without interception.

The application of electrostatic forces in fibrous filter filtration is a novel way to enhance filter efficiency. As mentioned in Professor Löffler's paper, there are several forms of electrostatic forces depending upon whether the particles, the fibers, or both of them are charged or whether an external electric field is applied. In the paper, Professor Löffler showed some interesting effects of coulomb force on filtration (i.e., both particles and fibers are charged), such as nonhomogenous distribution of particles behind the fiber. On the other hand, it also raises some questions which need to be answered. In the case of charged particles with fibers either charged or uncharged, the electrostatic force (or electric charge parameter) will change with time as charged particles continue to be collected on the fiber. What the effects of this would be are as yet unanswered. According to Fig. 6, his theory predicted that, for Stokes number < 1, the physical size of the particle is immaterial (i.e., independent of the interception parameter) and this needs to be checked experimentally using aerosols of different densities. In experiments where both particles and fibers were discharged, he showed that for paraffin wax particles (Fig. 10 in the paper), the Reynolds number had practically no influence between Re = 0.63 to Re = 2.63, while for NaCl particles Fig. 9), the Reynolds number clearly showed influence on deposition efficiency, even for Stokes number < 1 where rebound is not a problem. The reason for this discrepancy is unknown. Also it would be interesting to compare Figs. 9 and 10 of the data with the theoretical prediction shown in Fig. 3 and 4.

The coefficient of adhesion (sometimes called accommodation coefficient) is not only important in high velocity filtration,it is also important in sampling and/or sizing devices such as impactors. Professor Löffler's experimental approach, direct observation of particle rebound using a high-speed cinematography technique, would also be useful in investigating rebound problems in other devices. The probability of adhesion in filtration is an area where further studies are needed.

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OPEN DISCUSSION

Löffler: Thank you for the interesting comments,Dr. Yeh. You mentioned that the Kuwabara flow field in the experiment of Fuchs, confirmed the validity of Kuwabara flow field. We should not forget that Kuwarbara flow field assumed that the Reynolds number is very near to zero. If you deal with flow fields with Reynolds numbers between 4 and 30, I doubt if this is possible. Because we suppose that these Kuwabara flow fields are tested by Fuchs in a very low

Reynolds number range (about .001) and is only tested by a measure of direct force. The direct force with a velocity and pressure distribution in the very near neighborhood of the fibers is important. As for particle collection we need the velocity distributions in the whole flow field.

Perhaps I should point out that I talked a lot about theories. This is in no good accordance with the time we spent for experiments. We have spent a very short time, about 1 1/2 years, with the theories, and about 10 years doing experiments. That means we believe the theories are very helpful but we can not find a solution to the problem of filtration only from the theoretical point of view. We need experiments, especially since in industrial filters the situation is quite different from the theoretical models. That is why we do not like to emphasize the theoretical views. We should look more into the material behavior.

You mentioned there is a difference between the influence of Reynolds number to collection efficiency between parafin wax particles and filter. That is just the point. What I meant was the material behavior. For sodium chloride particles, we are sure to have an elastic rebound. Because we could recalculate adhesion probability, we could theorize the same behavior from the probability of adhesion. We measured the beginning of rebound of a particle with a velocity in the range of 5 to 7 centimeters/second. In our experiments with sodium chloride particles, and multifilter the lowest velocity has been 25 centimeters/second. It is in a range where adhesion probability is about 80%. This is our explanation for the difference between sodium chloride particles and wax particles.

One very important change is the electrostatic behavior of particles at the fiber. From the theoretical point of view, I can only expect there must be something going on during the build-up. We do not have a theoretical solution. We have tried to find experimental solutions:

ELECTROFLUIDIZED BEDS FOR INDUSTRIAL SCALE AIR POLLUTION CONTROL

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ABSTRACT

A review is given of the fundamental and practical development of the electrofluidized bed, beginning with application to the control of submicron particualte. Models and corroborating experiments establish fundamental particle collection coefficients, extend two-phase fluidized bed theory to account for bubble bypassing and describe the performance with ac excitations. Practical developments in the control of oil ash and asphaltic fume are described. Bed particle scale electromechanics are elucidated by recently developed models that are the basis for current efforts to apply the bed to controlling coal ash.

INTRODUCTION

As it is applied to air pollution control, an electrofluidized bed (EFB) is a shallow bed of particles, fluidized by the polluted gas, with an electric field applied by means of electrodes. Typically, the gas-entrained pollutant particles are electrically charged by ion impact prior to entering the EFB. Stressed as they are by the electric field, the bed particles act as collection sites for the pollutant. By design, they do not carry a significant net charge. The fluidized, usually semi-insulating, particles are subjected to an electric field that is imposed either transverse to the flow (the cross-flow configuration) or, by means of screen electrodes and the distributor plate, co-linear with the flow (co-flow configuration).

The following sections present an overview of fundamental and developmental efforts aimed at applying the EFB to large-scale air pollution control.

FUNDAMENTALS OF SUBMICRON PARTICULATE COLLECTION

A major incentive for using the EFB is the extremely short gas residence time required for collection of submicron particulate. A remarkably good model for plug-flow beds pictures the bed particles as being much like the electrodes of an ESP, with the electric field terminating over one side of each particle and originating on the other. Instead of the Deutsch model used for turbulent flow ESP's [1], the "local mixing model" is invoked with the assumption that on the average the fluid mechanics supplies the full surface of a bed particle with the gas-entrained particulate. The electric field simply brings the particulate from the local volume to the bed particle surface. According to this model [2], a bed having unfluidized height ℓ_o, bed particles of mean radius R, a superficial gas velocity U and an imposed electric field E collects particles of mobility b with the efficiency

$$\eta = 1 - \exp(-K_1) \; ; \; K_1 \equiv \frac{3\pi}{8} c \frac{\ell_o}{R} \left(\frac{bE}{U}\right) \qquad (1)$$

where c is a coefficient of order unity.

This model is valid provided that bubbling and reentrainment are not substantial. The first of these conditions is promoted by the several screens used to impose the electric field in the co-flow configuration. These tend to prevent the growth of bubbles. The second is not a problem for liquid aerosols such as DOP (used in fundamental studies).

With the objective of testing this model in a way that would separate out less fundamental (but perhaps essential) phenomena, extensive efficiency experiments have been carried out using an apparatus much like that shown in Fig. 1. For the data shown in Fig. 2, the cross-flow configuration shown in Fig. 1 is replaced by a co-flow bed. Efficiencies for DOP aerosols having two different mobilities are represented, the superficial velocity varies from 0.8 to 2.5 m/sec. For all of the measurements, ℓ_o = 6 cm and the bed particles are sand, R = 1 mm. The solid curve is Eq. (1) with c = 1.

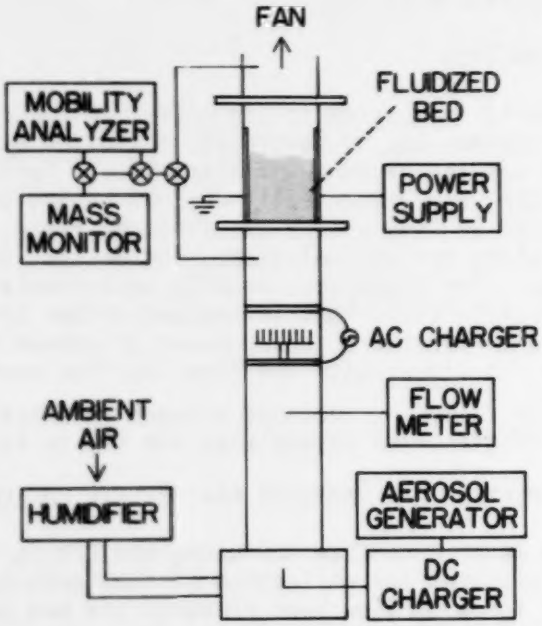

Fig 1 - Facility for fundamental room temperature tests using di-octyl phthalate aerosol.

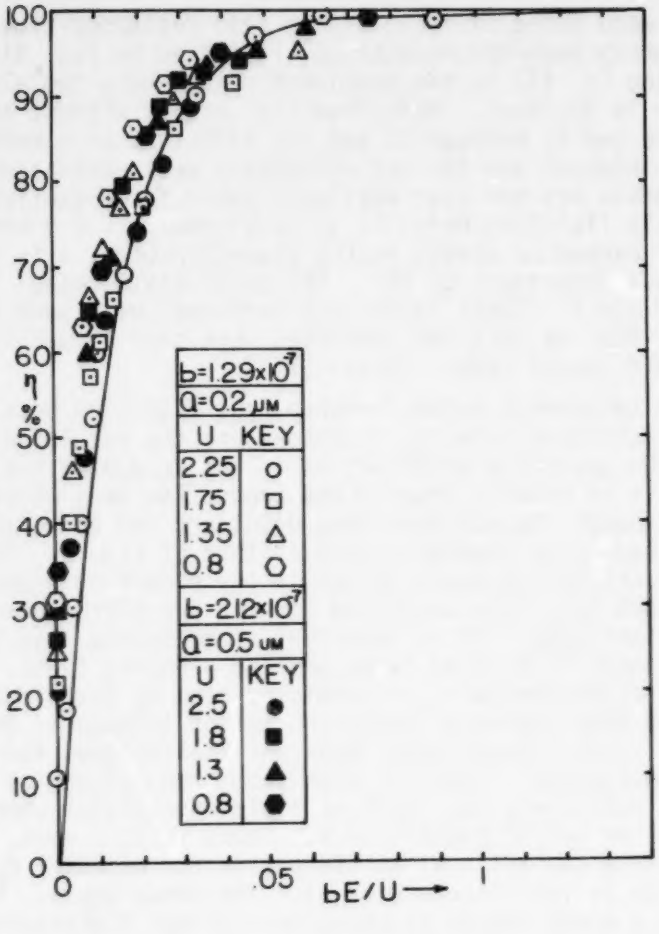

Fig. 2 - Co-flow efficiency as a function of bE/U for two different mobilities and sizes of particulate.

The residence-time advantage of the EFB is emphasized by recognizing that for an ESP having electrode spacing s, the Deutsch equation for the efficiency takes the form of Eq. (1), but with the collection coefficient K_1 replaced by $\ell bE/sU$. Because $3\pi c/8$ is essentially unity, this means that the unfluidized height of an EFB required to give the same efficiency as an ESP of length ℓ is $\ell_o = \ell(R/s)$. Typically, bed particles have a radius R on the order of 1 mm, so that for an electrode spacing of 10 cm the length of the EFB (not counting the charger) is theoretically 100 times less than that of the EFB.

Initial tests were aimed at establishing this residence-time advantage of the EFB. Subsequently reported results [2] (typified by Fig. 2) corroborated the model underlying Eq. (1) in the submicron range where the electrically induced collection is dominant. Note from Fig. 2 that without an applied electric field, the bed is between 20 and 30% efficient as a collector. With bubble bypassing minimized and the bed relatively well controlled in a plug-flow mode, these tests are the best available basis for establishing the parameter c. For highly fluidized beds, it is found that $c \approx 0.9$ with variations of about 15%. For packed or almost packed plug-flow beds, c is about unity and varies with test conditions by 25%. The local mixing model for dilute bed particles predicts $c = 1$. These tests were performed with sand at relative humidities of 50 - 90%, so that the particles were semi-insulating and hence the collection field placed under control [1,3].

Unless baffled to promote bubble breakup, gas fluidized beds do not expand uniformly as the fluidizing velocity increases and the resulting effects of nonuniformity in the gas-solid distribution is not accounted for in the plug-flow model. This is especially true in the cross-flow beds where the screen electrodes are not used. Models have been developed and correlated with experiments carried out in an apparatus such as that of Fig. 1. These combine the particle-scale collection model tested in the plug-flow experiments with two-phase models that have been developed for the transfer characteristics of conventional fluidized beds. It is important to recognize that the structure of beds can be strongly influenced by an applied electric field. In what will be termed the "bed electromechanics," particles tend to form "chains" or "strings" while gas flow causes a fluttering motion instead of the typical bubbling one [3]. These effects have important implications for bed mixing and particulate elutriation. However, high submicron collection efficiencies can be obtained at relatively low electric field intensities where electromechanical effects are not of significance. Thus, in this work, the Davidson model is adopted for a conventional bubble having the diameter D_b (in which the particle density is low) interacting with the dense phase. Part of the gas passes through a dense region at about the minimum fluidization velocity U_{mf}, while the remainder bypasses the system in the form of bubbles. There is a continual exchange of particles between phases accounted for by the Davidson model.

Two possible extremes can be used to model the penetration of particulate through the bed as a whole. The dense phase can again be viewed as evolving uniformly in the flow direction (the plug-flow model). Because of bubble agitation a complete mixing model is also plausible. Both models have been carried through in this study, with the dense phase plug-flow model clearly found to be more representative of what happens.

The plug-flow bubbling model shows how parameters not represented in the collection coefficient K_1 come into play. The predicted efficiency is

$$\eta = 1 - \frac{1}{m_2 - m_1}\left\{ m_1 e^{m_2 \ell_f}\left[1 + \frac{m_2 \ell_f U_{mf}}{c_q U}\right] - m_2 e^{m_1 \ell_f}\left[1 + \frac{m_1 \ell_f U_{mf}}{c_q U}\right]\right\} \quad (2)$$

where m_1 and m_2 are the roots of

$$m^2 + \left[\frac{(c_q + K_1)U}{\ell_f U_{mf}}\right] m + \frac{K_1 c_q U}{\ell_f^2 U_{mf}} = 0$$

and

$$c_q = \frac{9}{2}\,\frac{(\ell_f - \ell_{mf})}{D_b}\,\frac{U_{mf}}{U - U_{mf}}$$

The collection coefficient K_1 is the same as defined with Eq. (1), U and ℓ_f are respectively the superficial velocity and fluidized bed height under the conditions of interest while U_{mf} and ℓ_{mf} are these quantities under conditions of minimum fluidization. The most important parameter is D_b, the bubble diameter. The quantities necessary to evaluate these expressions are of course averages. The bed height is a fluctuating quantity and even in the shallow beds of interest here, bubbles grow significantly as they pass through the bed.

If the rate of gas interchange between the bubbles and the dense phase is large, which according to Davidson's model means that $c_q \to \infty$, the efficiency predicted by Eq. (2) reduces to that for the single-phase plug-flow model, Eq. (1). The most important aspect of the bubbling model is its prediction of efficiencies less than 100% no matter how large the collection parameter. In the limit $K_1 \to \infty$, the efficiency approaches the limiting value

$$\eta \to 1 - \left[1 - \frac{U_{mf}}{U}\right] e^{-c_q} \quad (3)$$

Using experiments to evaluate c_q, it is found that over a wide range of fluidization states, this parameter is mainly a function of the bubble diameter D_b.

In connection with cross-flow experiments without baffles, the two-phase model has been compared with extensive tests aimed at showing the dependence of collection efficiency on electric field intensity, bed particle size, static bed height, particle mobility and superficial gas velocity, with emphasis on the degree of bypassing through the bubble phase. The bed cross-section was 10 x 10 cm. In order to fix the electrical characteristics of the bed, the relative humidity was controlled at 90%. The fluidized particles were glass beads of 0.5 mm and 1 mm diameter as well as sand with mean diameters of 0.8 and 2 mm. The distributor plate was a 30 mesh plastic screen sandwiched between two insulating perforated plates with large holes.

Efficiency measurements over a wide range of parameters are summarized in Fig. 3, where the parameter is K_1 from Eq. (1). The solid curve is Eq. (1), and hence the limiting efficiency if bubbling is ignorable. A correction is made on the efficiency for mechanical effects. Unfluidized bed heights of ℓ_0 = 4,6 and 8 cm are represented by the data.

That the experimental efficiencies for these bubbling beds do not correlate with the parameter K_1 is evident from Fig. 3. In effect, the bed height becomes an additional parameter in correlating experiments carried out under different conditions. The data for the 6 cm beds is summarized in Fig. 4, where the ratio of experimentally observed to theoretically predicted efficiencies is shown as a function of the collection coefficient. The theoretical efficiency is given by Eq. (2).

The tendency for the model to predict the observations best at high values of K_1 is also seen in results for the other two bed heights. Correlation with the model over the full range of K_1 is best for the deepest bed.

Details of this work on effects of bubble bypassing are given in a forthcoming article [3].

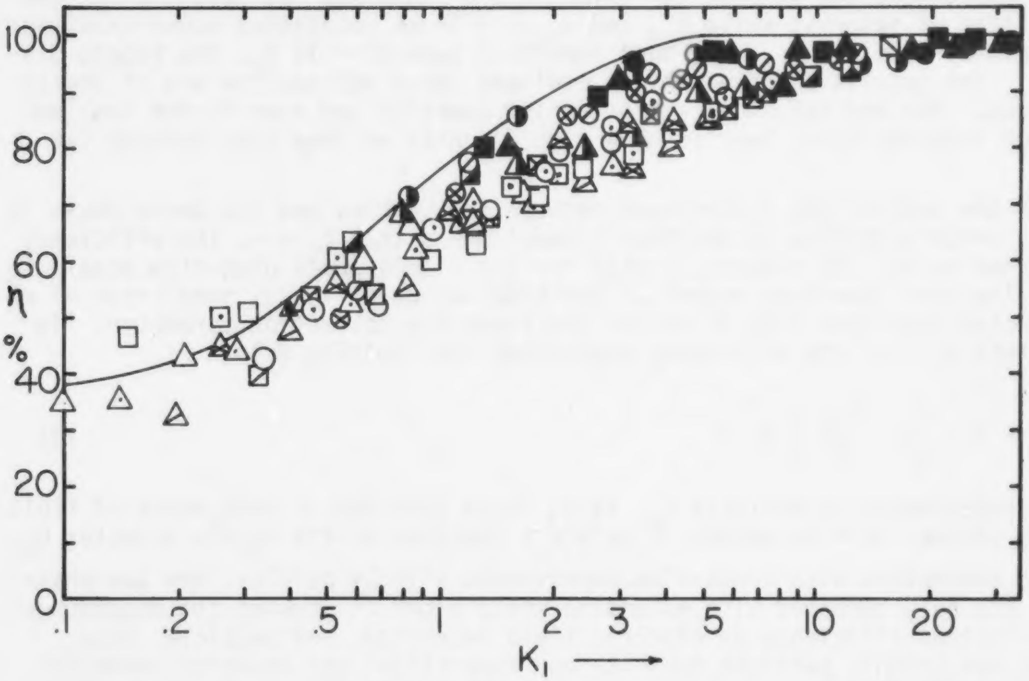

Fig. 3 - Composite of efficiencies for cross-flow beds as a function of collection parameter. Widely varying states of fluidization are represented.

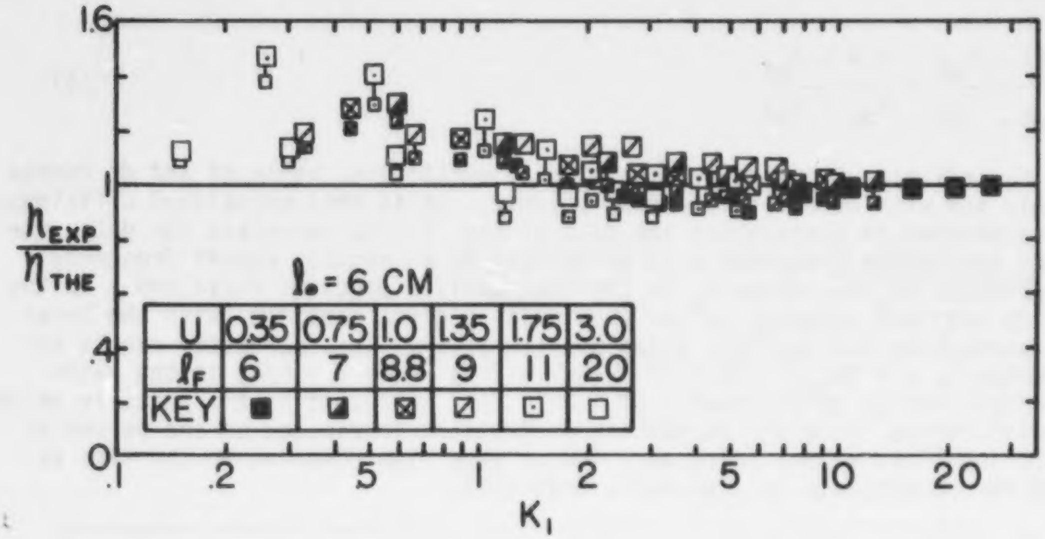

Fig. 4 - Measured efficiency normalized to theoretical efficiency as a function of collection parameter.

SUBMICRON PARTICULATES COLLECTION WITH AC EXCITATIONS

Savings from a reduced need for power conditioning equipment is an obvious advantage of being able to use alternating voltages to excite the EFB. But, perhaps a more important motive comes from the difficulty encountered when conventional ESP's are used to collect highly resistive materials. Poor performance is linked to the dc energization of the ESP. The collection of highly resistive ash results in the buildup of net charge and an attendant interference with the imposed precipitating field. Such effects are not expected in alternating field devices. To be effective with an ESP required to collect submicron particulate, alternating fields must have an extremely low frequency (on the order of 1 Hz). (Laminar flow ac ESP performance is worked out in detail by Alexander [5].) Can the EFB operate with 60 Hz excitations? More fundamentally, what are the implications of the frequency dependence of the efficiency for the basic collection process?

With beds of varying particle size and states of fluidization, and with applied fields of different amplitudes, experiments have been carried out using the apparatus of Fig. 2. Efficiencies were measured as a function of frequency. Both cross-flow and co-flow configurations were tested over the frequency range of 30 - 5000 Hz.

A correlation of the cross-flow data for the former experiments is given in Fig. 5. In order to emphasize the particle-scale collection process to the exclusion of bubble bypassing or bed scale mixing, the beds were tested at low overall efficiencies, in a range where the efficiency is proportional to the single particle rate of collection Γ. Thus,

$$\frac{\langle\Gamma\rangle - \Gamma_{nf}}{\Gamma_{dc} - \Gamma_{nf}} = \frac{\langle\eta\rangle - \eta_{nf}}{\eta_{dc} - \eta_{nf}} \qquad (4)$$

where <> indicates the time-average with ac excitation, while nf and dc denote no field and with dc excitation respectively. It is this normalized efficiency that is plotted in correlating the data of Fig. 5. To correlate the data, the angular excitation frequency ω is normalized to an angular cutoff frequency $\omega_e = \sqrt{9\pi/8(cs\, bE_o/R)}$, where E_o is the peak applied electric field and s is the velocity gradient adjacent to the equator of a given particle (with the local flow approaching the particle along a polar axis). This parameter can be approximated by $s \cong 3U_{mf}(1-K^5)/R\ (2-3K+3K^5 - 2K^6)$ where $K = R/R_o$ is the ratio of particle radius to the radius of a unit cell representing the particle voidage. The solid curves represent asymptotic predictions predicated on the period of the applied field either being very low or very high compared to the time required for particulate to traverse a unit cell.

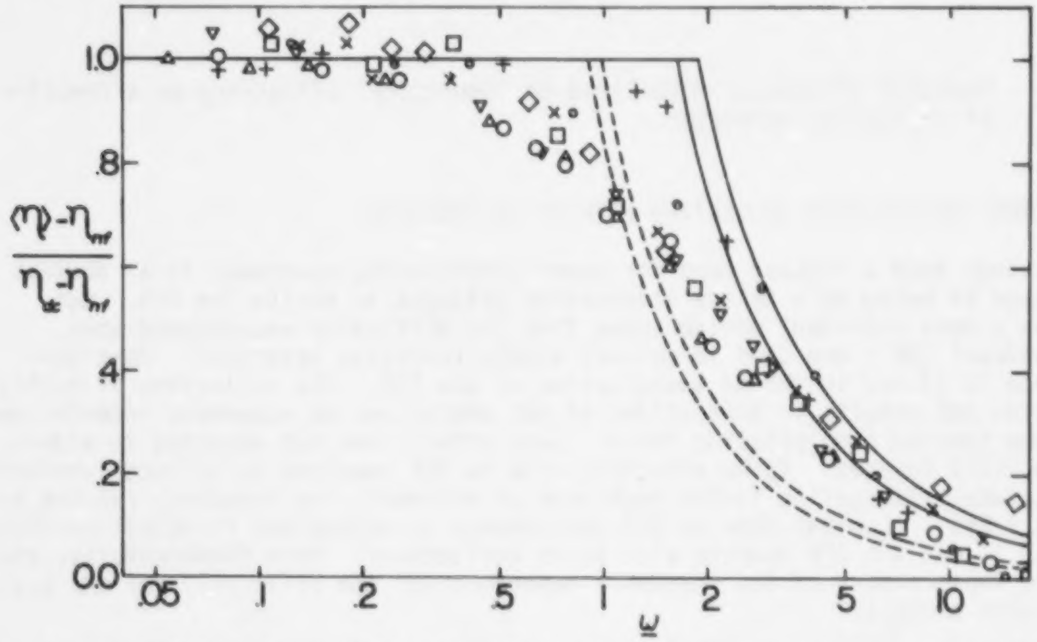

Fig. 5 - Normalized cross-flow EFB collection efficiency as a function of normalized frequency with low and high frequency theoretical asymptotes. Solid lines-cross-flow theory, dashed lines-co-flow theory. Rightmost of these curves for $\omega_e/\omega_m = 1$, leftmost for $\omega_e/\omega_m = 0.75$.

Details of the asymptotic models and specifics of the experiments will be available in the literature in the near future [6]. It seems clear that such frequency response measurements are a way of probing the mass transfer process at the bed particle scale. From a practical point of view, it is the fact that the 60 Hz efficiency is well below the critical frequency in all of the tests that is most significant. In this low frequency range, the efficiency is as predicted by the models for the dc excitations with $E_{dc} \to 2E_o/\pi$. As the field reverses, collection occurs first on one side and then on the other side of a collection site. The effective electric field is therefore the average of $E_0|\cos \omega t|$ over a full period.

Experiments, correlated with theoretical models, have been carried forward to understand how ac excitations can be used in the corona, ion-impact charging section. The general disposition of the charger just below the distributor plate is as in Fig. 1. At some penalty of power required, the charging can be accomplished at 60 Hz. This makes it possible to operate the entire EFB system, the bed and the charger, from a transformer. As an example, Fig. 6 shows the efficiency with the bed in this entirely ac excited type of operation. Details are given elsewhere [5].

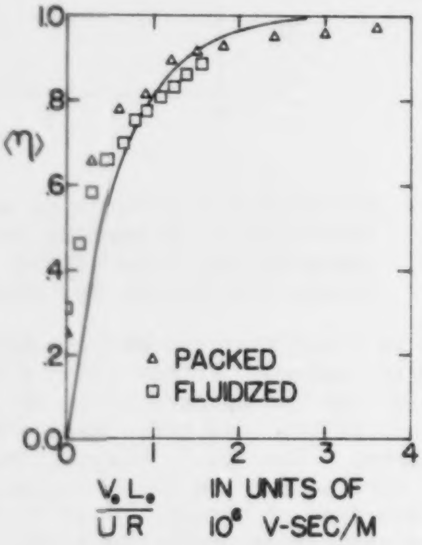

Fig. 6 - Efficiency with 60 Hz on EFB and charger. R = 0.5 mm, b = 1.2 x 10^{-7} (m/sec)/(volt/m). Fluidized: ℓ_o = 10 cm, U = 1.3 m/sec. Packed: ℓ_o = 7.5 cm, U = 0.5 m/sec.

COLLECTION OF OIL ASH

A natural application of the EFB for large-scale gas cleanup is to the products of oil combustion. Experiments have been conducted on a 100 cfm by-pass on flue gas prior to the economizer in the M.I.T. Central Utility Plant. Loss of heat in the connecting pipes cooled the gas to about 350° F. Tests were conducted in both the co-flow and cross-flow configurations using 1 mm glass beads and 0.8 and 2 mm sand.

The oil ash was entirely in the submicron range with loadings of the order of 0.01 to 0.06 grains/scf. Unlike the DOP, the oil ash is a dry material. Could the EFB be used to collect an essentially dry material and what residence time for the particles was consistent with retaining a reasonable efficiency? A typical test of efficiency as a function of time is shown in Fig. 7. For this case, a medium state of fluidization prevailed and the configuration was co-flow. The open circles represent collection with no charging and no applied field. Because the bed was at a sufficiently high temperature that the humidity did not render the particles semi-insulating, there were appreciable "micro-fields" associated with patches of charge on the particles created by frictional effects. Thus, the open half circles which represent data for collection with charging but without a field applied to the bed, show a considerably higher efficiency than found in experiments such as represented by Fig. 2. The solid data points are with both charger and applied field on.

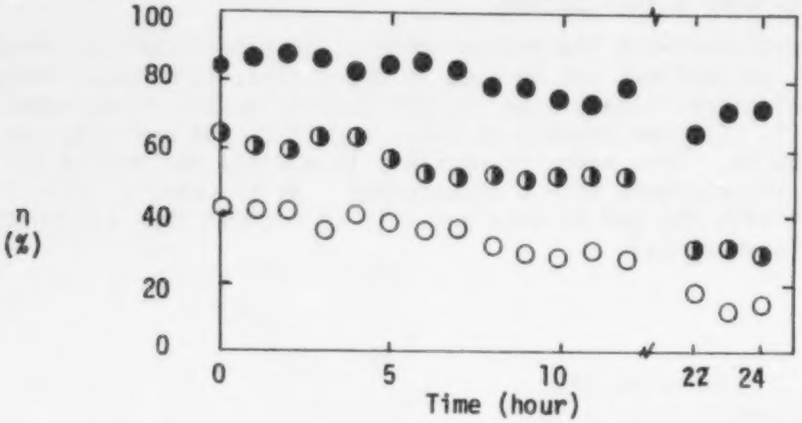

Fig. 7 - Efficiency as a function of time for co-flow EFB on bypass from M.I.T. Utilities while burning low sulfur oil. Open data points are for no-charging and no bed field. Half circles are with charging but no field. Closed circles are for charger and field on.

It is clear that the bed can operate for substantial periods of time at this mass loading. Either using a once through or a reprocessing system for the sand, the EFB appears to be an economically feasible approach to a utilities scale operation. However, there are questions that must be confronted by a development. The M.I.T. plant is required to use low sulfur oil and considerable excess air, so for the tests conducted the combustion products were extremely light and free of carbon. What is the effect on bed electrical losses of collecting combustion products of high sulfur oil? Losses with the tests reported here were insignificant.

A sketch of a co-flow EFB incorporated into a stack installation is shown in Fig. 8. The top view shows, to scale, the cross-section of a stack, with the EFB constructed around its base. Flue gas is fed into the annular EFB through eight ducts, passed through the bed and then into the stack through ducts in the side walls. The dimensions are for a 10^6 cfm unit. A typical pressure drop for 90 - 95% efficiency would be 10 cm H_2O.

CONTROL OF ASPHALTIC FUMES IN REPROCESSING OF ASPHALTIC HIGHWAY

Recycling of asphaltic pavement is being developed by the Warren Brothers Co. To be economical, the process should be carried out in adapted conventional plants. A major impediment to the conversion of existing plants is the formation of fine particulate smoke when crushed pavement is processed through the conventional aggregate drier. In the drier, the residual asphalt in the old crushed pavement begins to crack and release hydrocarbon vapor which subsequently condenses into droplets.

Devices currently available for collection of this submicron material are ill suited. The asphaltic by-products are not easily removed from the plates of an ESP. Fabric filters that will catch the submicron particulate subsequently

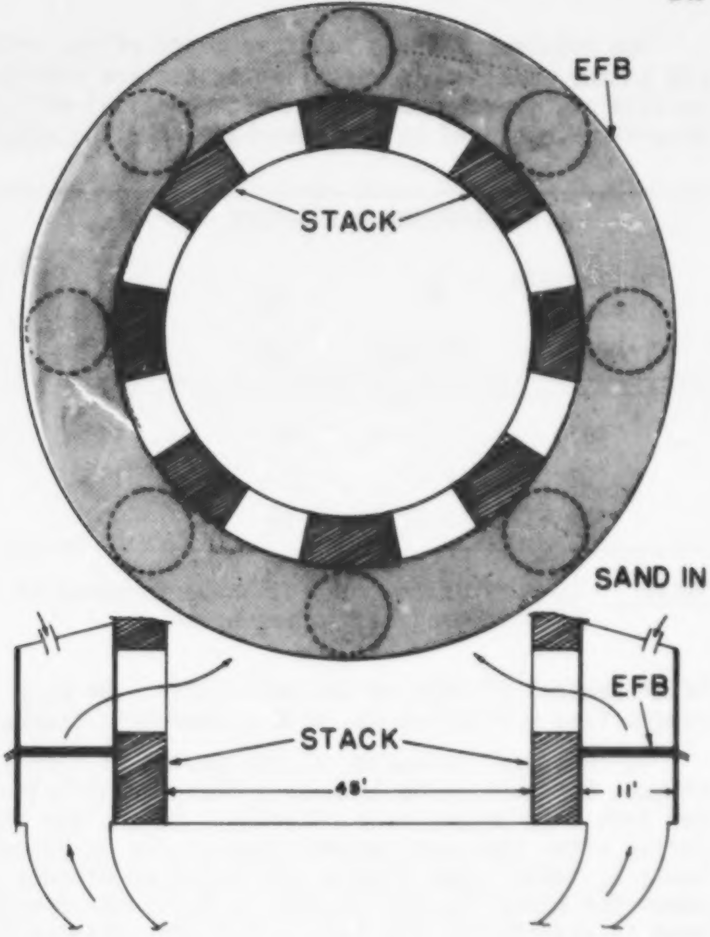

Fig. 8 - A 10^6 cfm EFB constructed around the base of a stack.

are fouled. Wet wall electrostatic precipitators, charged droplet scrubbers and high energy Venturi scrubbers all create a water pollution problem. The EFB is well suited to this application, not only because of its high efficiency in the submicron range, but because the material used to collect the particulate is easily processed through the EFB and conveniently incorporated into the asphalt hot mix product. In fact, in this and other related operations oiling of the sand is a desirable step in the plant process.

A prototype EFB of cross-sectional area 1 ft^2 has been tested on both a small-scale prototype asphalt recycling system and on a bypass to a full-scale asphalt plant operating in the recycle mode (Warren Brothers plant #135, Richmond, Va.). In both of the tests reported here, a bag-house was used ahead of the EFB to catch large dust particles from the drier. The submicron material easily penetrated this bag-house. Tests were performed on a batch basis using sand of 2 mm median diameter. The charger was of the plate wire type. Temperatures ranged up to 500° F.

The results of three separate tests of the collection efficiency of the EFB on the small-scale recycling system are reported in Table 1. Experimental results are shown along with the theoretical efficiencies predicted by the plug-flow model and by the two-phase bubbling model.

EFB OPERATING PARAMETERS				EFFICIENCY (%)		
				Experimental	Theoretical	
Particulate Loading (mg/sm^3)	U (m/sec)	ℓ_0 (cm)	E (kv/m)		Plug-flow Model	Bubbling Model
167	1.7	10	465	96	99	96
82	1.8	10	250	93	94	91
44	1.8	13	500	94	99	92

Table 1 Summary of the collection performance of the EFB; comparison of experimental and predicted efficiencies.

After removal of dust by the bag-filter, the mass median diameter of the smoke ranges from 0.3 to 0.5 µm, with a geometric standard deviation of about 2.

An important parameter is the amount of time that sand can be used before it must be replaced because of a drop in collection efficiency. The mechanism for this drop seems to be channeling of the gas through spouts in the bed, occuring after the sand becomes viscous due to excessive oil collection. In Table 2, tests under similar operating conditions are compared. In the two cases for which the oil loading is 0.7%, the measured efficiency for the oily sand is essentially the same as for the clean sand. In the case of 1.2% oil loading, a drop in efficiency is observed.

EFB OPERATING PARAMETERS			EFFICIENCY (%)				
			Fresh Bed		Oil Coated Bed		
U (m/sec)	E (kv/m)	ℓ_0 (cm)	Particulate Loading (mg/sm^3)	Experimental	Particulate Loading	Oil Loading *	Experimental
1.8	500	13	44	94	107	0.7%	94
1.8	500	13	44	94	200	0.7%	95
1.7	465	10	167	96	233	1.2%	85

* Oil loading is defined as the ratio of the weight of collected oil to the bed total weight.

Table 2 Comparison of collection performance of fresh and oil coated beds.

Pressure drop usually accounts for the largest share of EFB operating costs. In these tests, the pressure drop across the EFB, including the distributor plate, was less than 18 cm H_2O. With the bed operating properly, this drop is essentially that across the distributor plate plus what drop is required to support the bed. As spouting occurs because of bed overloading, there is an associated loss of pressure drop that can be taken as an external indication of loss in efficiency.

Results from the Richmond tests are shown in Table 3, where 400 cfm is filtered with particulate loadings varying between 4 mg/m^3 and 41 mg/m^3. Particulate was observed to have a mass median diameter of 0.35 μm with a geometric standard deviation of 1.4

Efficiency Measuring Device	EFB Operating Parameters			Efficiency
	E(kv/m)	U(m/sec)	ℓ_0(cm)	(%)
Mass Monitor	400	1.9	10	99.3
Andersen Impactor	400	2.1	11	99.1

Table 3 Summary of collection performance in Richmond.

This work will be reported in more detail in a forthcoming article [8].

ELECTROMECHANICS OF ELECTROFLUIDIZED BEDS

In the application of the EFB to submicron particulates control, it is possible to separate the electrical interaction between the bed particles and the charged particulate from the electrical interaction between bed particles. It is the latter that is termed the bed "electromechanics." The propensity of the field for altering the state of fluidization by inducing interparticle stringing is discussed in previous publications [3].

If the bed is to be applied to collection of both submicron and supermicron particles, it is necessary to understand the implications of the surprisingly large electrical effects at the bed particle scale. To what extent can these forces be relied upon to prevent loss of large particles from the bed? Because the field can turn a bubble into a "fritter," the turbulence within the bed is considerably reduced by the field. Each bed particle is subjected to a reduced amount of agitation because the "wake mixing" associated with the classical bubble is reduced or eliminated. To what extent are these effects useful in achieving EFB operations in a mode where bed particles are agglomerates of the particulate, the self-agglomerative mode?

Two types of fundamental experiments have been carried out with the objective of having a physical picture of the interparticle force. In the first of these, the EFB (in a cross-flow configuration) is in the annulus between the cylinders of a Couette viscometer, as shown in Fig. 9. The shear stress

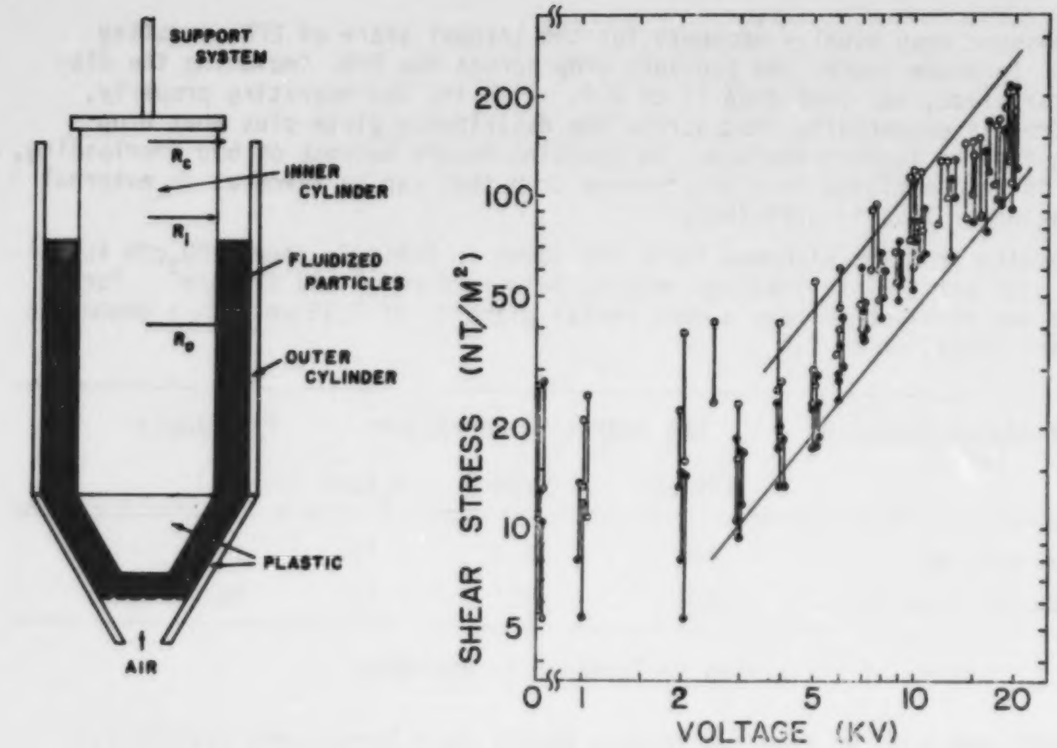

Fig. 9 - Cross-flow EFB in viscometer. Outer cylinder rotates and inner one is fixed. Voltage applied between inner and outer cylinders results in shear stress at inner cylinder as shown.

tends to a common asymptote at high field strengths over a range of shear rates. In this high field regime, the electrical force is large enough that the bed is electropacked. Each particle chain transmits its force through friction to the electrodes, so the shear stress can be used as a measure of the interparticle force.

For the particles typical of the EFB applied to air pollution control, polarization forces (associated with the permittivity of the particles) are too small to be responsible for the observed effects. Also, such observations as the dependence of the electromechanical effects on applied voltage negate such polarization forces as the basis for what is observed. A theoretical model has been developed in which the force between particles is accounted for in terms of the concentration of the electric field near the contact caps between particles. In this region there is a strong constriction of the current as it passes between particles. The correlation between theory and experiment supports the contention that it is this field concentration that is responsible for the large electromechanical effects and their dependence on the applied voltage. This work is reported in more detail in a forthcoming article [9].

Further experiments allow deduction of this interparticle force from bed characteristics. In the first of two experiments in this category, the pressure drop at incipient fluidization is measured as a function of the applied voltage. In the second, the minimum voltage required to suspend a bed against an upper screen is measured as a function of the gas-flow rate. Both experiments are performed using the semi-insulating particles typical of the sand at high relative humidities with the cross-flow configuration. Typical results for this latter experiment are shown in Fig. 10, which shows the superficial velocity at which particles fall from a pinned position as a function of voltage. The inset shows an idealization of the curves, identifying two regimes. In the low-field regime, the velocity is above U_{mf}, and so there is enough pressure drop to support the bed as a whole. In this regime, failure is because particles at the bottom interface are not held in equilibrium. In the high-field regime, the electric field makes it possible to support the bed as a whole and failure involves the entire bed. Correlations between the bed measurements and a theory based on interparticle forces due to current constriction and a contact cap determined in size by electrical breakdown are not only successful in predicting the dependence of the parameters, but in giving quantitative predictions as well. This work will be reported in the literature [10].

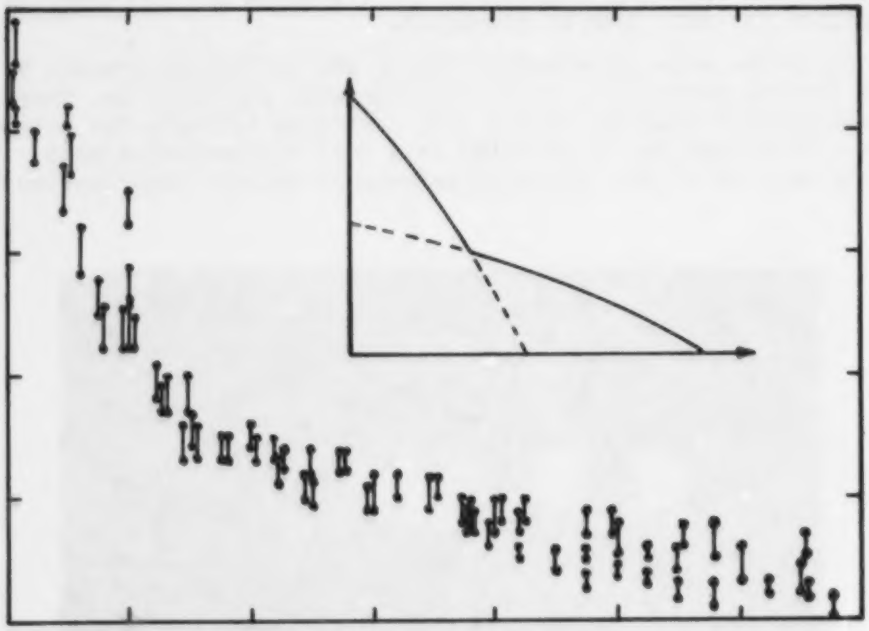

Fig. 10 - Air velocity required to support a cross-flow electropacked bed against an upper stream as a function of the applied voltage.

COLLECTION OF COAL ASH

Work is being carried forward by Alexander (Ph.D. research) to apply the EFB to the control of coal ash. Experiments are on a 100 cfm facility that ranges to 600° F in temperature. Two approaches are being investigated. In the first, the bed particles are coal, which is used as a collection site for ash in a process that builds up particles of about twice the original mass before the coal and ash particle is removed from the system. Combustible additives are used to promote stable agglomeration. A picture of one of these composite products from the EFB is shown in Fig. 11. A typical collection efficiency for the bed is shown in Fig. 12, where the advantage of using the EFB for the submicron collection is apparent. Note that a fluidized bed without electrification can be used for removal of particles above about 3 μm. In the system being developed, two beds are used. In the first, the large particles are removed and in the second, charging and application of a field results in good efficiency for collection of the fines.

In the system using coal as the seed material, the output of the bed is to be injected into the combustor where the ash cracks away and the coal and additive are burned. Provided that the combustor is one of relatively high ash retention, this approach can be shown to lead to good overall efficiency with the ash removed from the combustor. The fluidized bed combustor is a natural candidate for this mode of operation.

The second system under development starts the collection process by "seeding" the bed with foreign particles. As the agglomerates are built up, they are removed, treated, and a fraction broken into particles suitable for reinjection into the bed. Thus, the bed is operated in a self-agglomerative mode. Detailed studies are being made of the interplay between parameters contributing to this process.

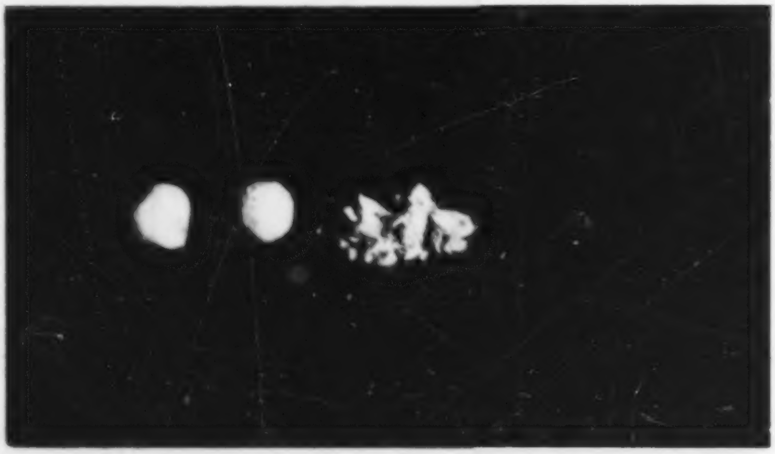

Fig. 11 - Composite ash and coal particles built up in the EFB. Coal particle at right (2 mm diameter) is seed for agglomerates shown at left. The third particle from the left has been crushed.

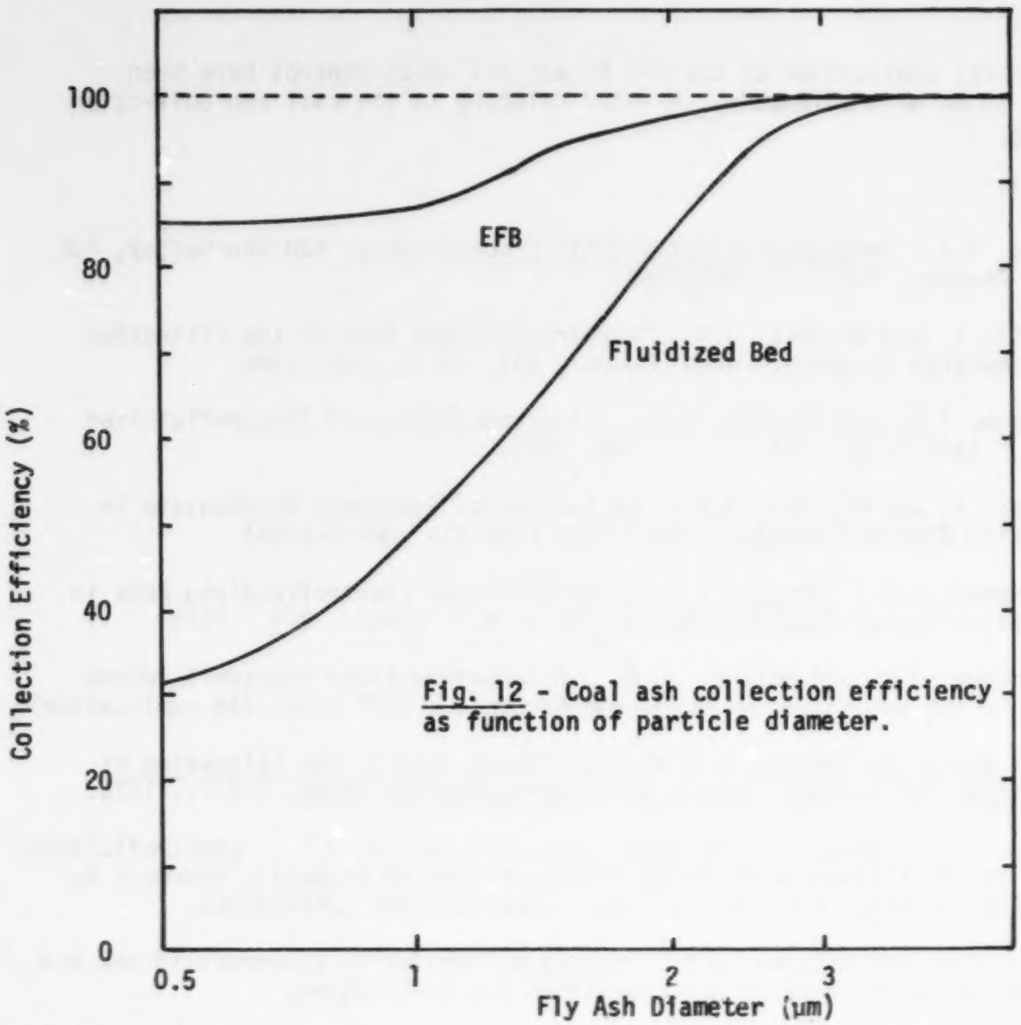

Fig. 12 - Coal ash collection efficiency as function of particle diameter.

SUMMARY REMARKS

It should be clear from the current developments that the use of additives would improve the viability of the EFB in the oil ash application. Indeed the range of applications, not only to particulates control but to chemical processing in general is rapidly expanding.

ACKNOWLEDGMENTS

The tests on asphaltic fume control were supported by the Warren Brothers Co. in collaboration with Mr. J.F. Denton. Otherwise research reported here was sponsored by the Empire State Electric Energy Research Corporation. Contributions by the authors,colleagues and students are recognized from the references, which are based on M.S. and Ph.D. theses by K. Zahedi, a M.S. thesis by J.C. Alexander, a B.S. thesis by P.B. Zieve and a Ph.D. thesis by P.W. Dietz. Work on coal ash is being carried out by J.C. Alexander. E.P. Warren made experimental contributions. Two patents held by M.I.T. and relating

to the general application of the EFB to air pollution control have been indicated to be held allowable. A third relating to the coal ash collection is pending.

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WRITTEN DISCUSSION

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My experience with electrofluidized beds differs somewhat from Jim Melcher's experience in that in my work the beds were used to supply powder to electrostatic coating and printing processes which required moving particles larger than 10 microns through or out of the beds. Particles finer than 10 microns and especially finer than 1 micron were avoided because the surface molecular forces or Van der Waal forces cause the fine particles to stick to the larger bed particles.

I have been very interested in Jim Melcher's work and have been following it for some time. Although a number of questions can be raised on the basis of the information mentioned in this paper, these questions have been answered in papers which have been published earlier. I strongly recommend these papers for further reading to anyone interested in electrofluidized beds.

I find a problem with only one part of the paper. The statement is made that "Typically, bed particles have a radius R on the order of 1 mm, so that for an electrode spacing of 10 cm the length of the EFB (not counting the charger) is theoretically 100 times less than that of the ESP." The factor "100" was obtained by equating the K coefficients in the Deutsch equation and in the efficiency equation of the electrofluidized bed. A factor of 2 was omitted from the K factor of the Deutsch equation. Thus, the length of the equivalent ESP should be 200 times rather than 100 times longer than the length of the EFB.

Since I have no further comment about this work, I shall spend the rest of the alloted time in discussing another new air cleaning system which is called an electroinertial air cleaner.

Inertial air cleaners consisting of a bank of straight flow tubes, each tube having a swirl means to cause dust to be centrifugally thrown outwardly toward the tube wall, are well known for engine air cleaning application. Dust near the tube wall is drawn from the main stream by a scavenger fan and cleaned air emerges from the core zone of the tube through a small diameter take-off tube extending into the downstream end of the flow tube. However, the centrifugal separator action is relatively ineffective on particles smaller than 5 microns; therefore, electrostatic separator action was added in electroinertial devices to enhance overall collection efficiency. In one arrangement, a fine wire at a potential of approximately 15-20 kV is extended through a 1-1/2-inch-diameter tube on the tube centerline with the tube wall at ground potential to establish a flow of negative ions. The resultant negative charges on the particles and the radial electrical field from the ionizer wire to the tube enhance outward migration especially of the smaller particles thereby improving overall collection efficiency.

Unfortunately, the electrostatic separator action causes some particles to precipitate and adhere rather strongly to the tube side wall. Therefore, the tube usually must be periodically rapped or vibrated in the radial and/or axial direction in order to dislodge the particles sufficiently to permit the scavenger air to carry them away. A rapper equipped tube is shown in Figure 1. Under some circumstances the particles are jarred with such force as to be reentrained into the clean air stream. Even when properly applied, rapping or vibrating requires special shock mounting of the tube; mechanical wear on the mounts is a problem. Therefore, the use of rappers as a dislodging expedient

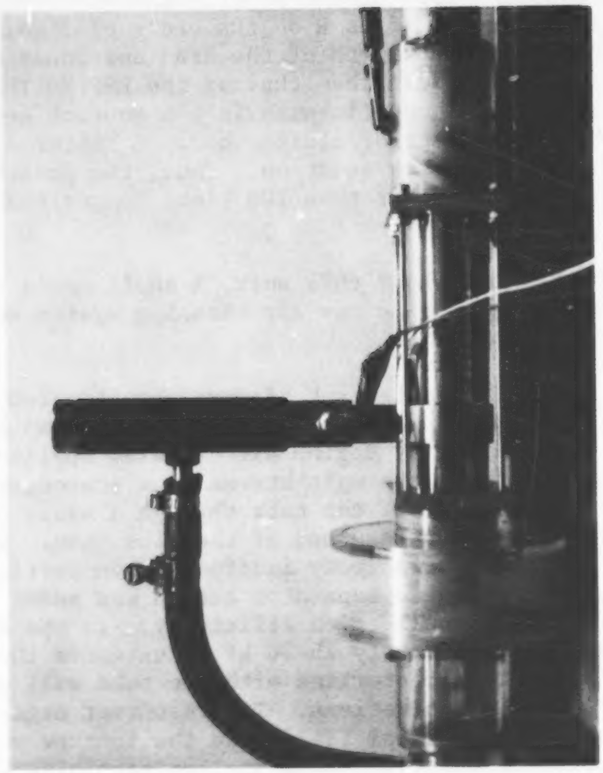

FIGURE 1. ELECTROINERTIAL UNIT WITH RAPPER

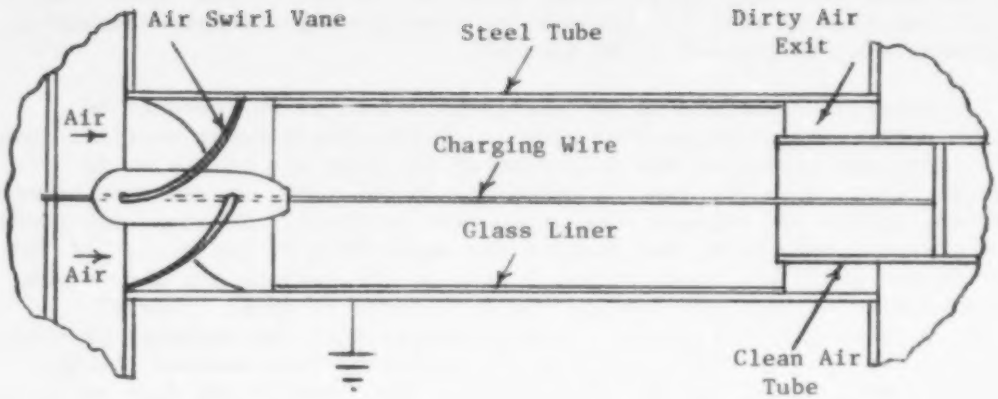
Air Swirl Vane
Steel Tube
Dirty Air Exit
Air
Charging Wire
Glass Liner
Air
Clean Air Tube

FIGURE 2. ELECTROINERTIAL UNIT WITH GLASS LINED TUBE

is not entirely satisfactory.

While working with a unit of this type under Contract No. DAAE07-72-C-0293*, a glass lining was tried in the tube as shown in Figure 2. During trials without vibrating the tube,AC fine test dust was passed through a 1.5-inch-diameter by 6-inch unit at a flow rate of 40 cfm including 4 cfm of scavenger air. No measurable amount of dust accumulated in the tube. The fact that no dust accumulated in the glass-lined tube represents a major advance in the technology. The overall efficiency was about 92 percent. For comparison, with a vibrated unlined tube with comparable voltage on the wire, the overall efficiency is generally about 99 percent. As an inertial unit, with no applied voltage, the same size unit would remove about 80 percent of the dust. The electrical contribution with the glass-lined tube was significant--increasing the efficiency from 80 percent to 92 percent.

The low material retention in the glass-lined tube is believed to be unique to "leaky" insulators such as glass. In the precipitator section, large numbers of ions are generated but only part of the ions are effective in charging the dust. Once the dust is deposited on the tube, excess ions bombard the dust and inhibit its release from conductive surfaces. However, the excess ions accumulate on insulators and prevent the deposition of the dust. If the insulator is too good, the potential will rise on the surface and the corona from the wire will be shut off causing the efficiency to drop. "Leaky" insulators, such as glass, let some current pass so that the charging process continues but some charge accumulates on the surface. This surface charge, which is the same polarity as the charge on the dust, repels the dust as it approaches and prevents it from making good contact with the surface. As a result,the surface molecular forces, which act over very short distances and cause fine particles to adhere tenaciously when good contact is made, are not effective. The loosely held dust thus moves along the charged surface of the tube and out through the scavenger system.

Recently U.S.patent 4,010,011 was issued on this device and further work has been done for the USDA on other modifications of the basic electroinertial design for removal of fine cotton dust.

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*USA TACOM.

OPEN DISCUSSION

Oder: The question has to do with bubbles and the suppression of bubbles with the electrofluidized bed. What is the fundamental mechanism for the creation of bubbles and why should the electrostatic field suppress that?

Melcher: The basic reason for the bubbles, is instability in fluid. Usually that is how it is approached in the literature of fluidized beds. It is unstable, and the nonlinear equilibrium that results is the bubble. The other question, how is it that the electrostatic field seems to fill in the bubble? Again this is the sort of dependency in this self precipitation time constant. Particles do not have net charge, they do not go off the wall, and they do not annihilate each other. But they are polarized and that now has a new time constant. That time constant is now in the steady state configuration in competition with the turbulence and, at a certain point, is short enough that it overcomes the inherent mixing time constant and they form strings. These strings are held together by this capped force which we are talking about. I can use an electric field to support a static bed very deep in spite of the fact that a field that large will not support water at an altitude of a fraction of a millimeter. The force between the particles is a very, very strong one and it can be transmitted to the walls. That is why you get the overshoot. It can change the effective shear stresses as well. Students in industry are now working with SO_2 on the type of approach.

Oder: Let me further my question. You are not really getting rid of a fundamental instability. You have really put a gas distributor in there. In the bed where you have this string of particles do you have mass bed mixing or is the bed stable?

Melcher: The bed is indeed mixing. I have a movie with me if any one wants to see it. With the field on, the bed mixing is much slower. The way in which the particles are treated by the bubbles is different because the bubble goes through and you get strong mixing in the trail. What the field does is eliminate that trail, it eliminates a lot of the agitation in the bed.

Koppang: I was wondering what your mechanisms are for cleaning your bed.

Melcher: I appreciate your question. Two advantages of the thing are, first of all, the savings you get in the pressure drop and next, it flows like a fluid. You put a hole in this thing and it runs out. You can run this thing on a continuous basis. Generally you let it run down through a hole and you put a little in. It is a very slow rate at which you do that. They use more oil and sand than we can supply. For certain kinds of applications I can see for fly ash you might agglomerate and recycle the whole thing.

On the fly ash we have two approaches. One is to self agglomerate the fly ash on itself. The other is to agglomerate on coal. In either case we want to recover the heat value of additive of the coal injected into the combustor. A combustor like a fluidized bed combustor has a high ash retention, so that taking all the ash out of the combustor is the grand plan. Also I might mention we are working on recycling seed for MHD generators, where we have a local pollution problem in our MHD experiment. The sodium carbonate that we put in as seed turned out a submicron particulate all over the campus and an EPA warning. We've gotten ourselves a 70 inch pressure drop scrubber which gets 70% of the stuff. We are trying to collect it ourselves since it is valuable. It is a nice application; we can put it right back into the combustor.

APPROXIMATE EQUATIONS FOR PREDICTING ELECTROSTATIC PARTICLE COLLECTION

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ABSTRACT

A new approach to calculating approximate single-collector collection efficiencies is presented that eliminates the need to calculate particle trajectories, for particles of negligible inertia. The general case of a central force, $F = k/r^m$ $(m \neq 0)$, is solved for spherical and cylindrical collectors having collection efficiencies either much greater than or much less than one. For those values of m for which efficiencies have already been evaluated by others (using trajectory integrations), the results agree quite well, except for the low-efficiency case of a charged particle and an uncharged collector, where the central force approximation breaks down. The approach taken here emphasizes the significance of the aerosol concentration profile, n(r). It shows that increased collection efficiency will result from improved mixing of the aerosol in the vicinity of the collectors, perhaps through induced turbulence. The collection efficiencies are expressed in terms of $K = w_s/v_o$, the ratio of the particle radial velocity at the collector surface (due to the central force) to the mean gas velocity far upstream from the collector. The collection efficiencies are as follows:

Collector	$K << 1$	$K >> 1$
cylinder	πK	$(\pi K)^{1/m}$
sphere	$4K$	$(4K)^{2/m}$

The results are exact for coulombic attraction (charged particle, charged collector, point and/or line charge approximation), if inertia is negligible and if interception is not important enough to produce critical trajectories which strike the collector upstream from the rear stagnation point. Also discussed are the effects of other particles, other collectors, and other fields, with respect to the collection efficiency of a single collector.

INTRODUCTION

Recent emphasis on the control of respirable dust, airborne particles with aerodynamic diameters $\leq 3\ \mu m$, has led to increased interest in using electrostatic forces to enhance the collection in filters and scrubbers. To predict the enhanced collection efficiency of these control devices exactly is very difficult. Often one can determine whether or not the addition of electrostatic forces to a conventional scrubber or filter would be a major improvement through the use of approximate equations for collection efficiency.

GOALS

The goals of this work are to make it easier to calculate approximate collection efficiencies of control devices (such as scrubbers and filters) which employ electrostatics to augment their usual collection mechanisms and to clarify the role of electrostatics in such devices. The alternative to such simple calculations is often the rather difficult determination of particle trajectories for the variety of conditions and particle sizes of interest, an expensive endeavor.

ANALYSIS

A. Particle Motion

Denote the particle velocity by $\vec{u}$ and the gas velocity by $\vec{v}$. The well-known equation of motion for a particle governed by Stokes' law with the Cunningham slip correction is

$$m\, d\vec{u}/dt = -3\pi\mu\, d_p(\vec{u} - \vec{v})/C + \vec{F} \qquad (1)$$

in which m is particle mass (g), t is time (s), μ is gas viscosity (poise), d_p is particle diameter, C is the Cunningham slip correction (approximately $1 + 0.16\ \mu m/d_p$ at NTP), and $\vec{F}$ is the vector sum of any external forces acting on the particle (dynes). Fuchs[1] discussed this equation's range of application in detail. It is generally applicable to particles with diameters from 0.1 μm to 10 μm.

Equation (1) can be simplified by introducing particle mobility, B, given by

$$B = C/3\pi\mu d_p \qquad (2)$$

for spherical particles. The mobility is the terminal velocity of the particle with respect to the gas per unit applied force. The mobility is also the ratio of the particle aerodynamic characteristic time, τ, to the particle mass:

$$B = \tau/m \qquad (3)$$

where, for spheres,

$$\tau = C\rho_p d_p^2/18\mu \qquad (4)$$

(ρ_p is the particle density).

Equation (1) then becomes

$$\tau\, d\vec{u}/dt = -(\vec{u}-\vec{v}) + B\vec{F}. \qquad (5)$$

The first simplification we make is to eliminate the inertial term $\tau\, d\vec{u}/dt$. We distinguish between the "control device" and the "collectors" (fibers, droplets, etc.) within the device. Let the mean flow velocity (or the flow far upstream from the collectors) be v_o and let the collector's smallest dimension be L. Then $\tau\, d\vec{u}/dt$ becomes negligible when

$$v_o\tau/L << 1 \qquad (6a)$$

$$BF\tau/L << 1 \qquad (6b)$$

under which conditions

$$\vec{u} = \vec{v} + B\vec{F} \qquad (7)$$

Inequality (6a) means that the particle inertia is insufficient to cause it to deviate appreciably from the streamlines in the vicinity of the collector. Inequality (6b) means the particle inertia is insufficient to prevent the particle from almost instantaneously achieving its terminal velocity with respect to the fluid due to the force F. Equation (7) indicates that under these assumptions the particle motion is the motion of the gas plus the particle terminal velocity with respect to the gas, $\vec{w} = \vec{F}B$.

B. Particle Collection

Figure 1 shows a collector of arbitrary shape in an aerosol flow of free-stream velocity v_o and free-stream concentration n_o. The number of particles collected per unit time at steady-state will be the integral over the collector surface of the particle velocity normal to the surface times the concentration:

$$\dot{N} = -\oint_{\vec{A}_s} n_s \vec{u}_s \cdot d\vec{A}_s \qquad (8)$$

where $\vec{A}_s$ is the surface area (cm^2), $\vec{u}_s$ is the particle velocity (cm/s), and n_s is the particle concentration at the surface (cm^{-3}). In steady state, the number of particles removed per unit time from the volume (spherical or cylindrical) within R_2 is also $\dot{N}$,

$$\dot{N} = -\oint_{R_2} n_2 \vec{u}_2 \cdot d\vec{A}_2 \qquad (9)$$

Evaluation of the preceding integrals requires concentrations, velocities, and surface areas. One simplification is that at the surface of the collector the normal component of velocity for the gas must be zero, so Equation (8) becomes

$$\dot{N}_s = -\oint_{A_s} n_s \vec{w}_s \cdot d\vec{A}_s \qquad (10)$$

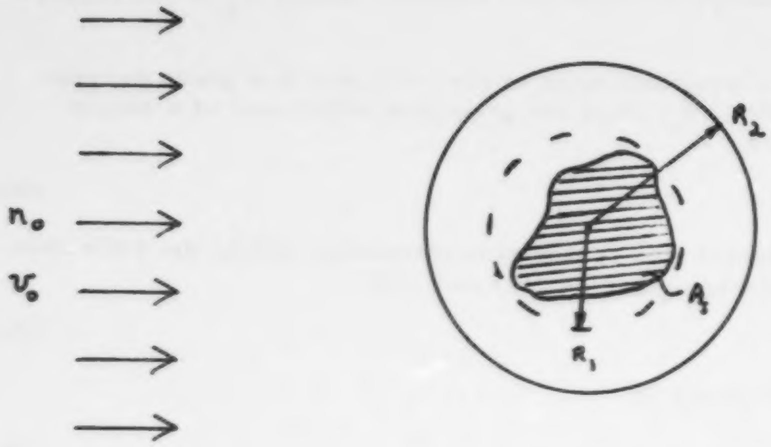

Fig. 1 Collector of arbitrary shape in aerosol flow of free-stream velocity v_o and concentration n_o.

C. Collection Efficiency - Coulombic Attraction Only

Coulombic attraction is the attraction between one point charge (the) particle) and another (the collector) or between a point charge and a line charge (a cylindrical collector). The charges are opposite in sign and are treated as being located at the centers of the particle and the collector. The force between the point charges decreases with the inverse square of the distance (r^{-2}); the force between the particle and the line charge decreases inversely with distance (r^{-1}). This force will often predominate over the other electrostatic forces. Equation (1) becomes easy to evaluate because, as Levin demonstrated (see Fuchs [1]): for forces, such as gravity or coulombic attraction, for which the divergence is zero; that is, for

$$\vec{\nabla} \cdot \vec{F} = 0 \qquad (11)$$

the concentration of particles in an inertialess aerosol subjected to that force is constant along particle trajectories. Thus, the surface concentration at the collector will be the same as the free stream concentration

$$n_s = n_o \qquad (12)$$

which means

$$\dot{N} = n_o w_s A_s \tag{13}$$

for a spherical or cylindrical collection surface, where A_s is the surface area.

If A_c is the cross-sectional area of the collector in a plane perpendicular to the mean flow, $\vec{v}_o$, then the collection efficiency of a single collector is defined as

$$\eta = \dot{N}/n_o v_o A_c \tag{14}$$

For a cylindrical collector and couloumbic attraction alone, the collection efficiency becomes, from equations (13) and (14)

$$\eta = \pi\, w_s/v_o \tag{15}$$

and the collection efficiency is

$$\eta = 4w_s/v_o \tag{16}$$

for a sphere. Usually, efficiency is expressed in terms of the ratio, K, of the electrostatic force on the particle at the collector surface (F_s) to the drag force (F_D) associated with a velocity v_o, but in fact

$$K = \frac{F_s}{F_D} = \frac{w_s/B}{v_o/B} = \frac{w_s}{v_o} \; . \tag{17}$$

Note that the coulombic force decreases as rapidly as the collection area increases; thus, the collection rate, $\dot{N}$, is independent of size, even though the collection efficiency is dependent on size. If particulate capture by interception can be correctly modeled as an increase in the effective collector size (increased outward from the surface by a dimension equal to the particle radius), interception will not increase the rate of capture, unless $w_s(r=R+r_p)$ at the collector rear is smaller than $v(r=R+r_p)$ there.

It is also significant that the rate of capture $n_o w_s A_s$, is independent of the flow profile, and so is the efficiency.

Further, if a surface such as R_1 in Figure 1 can be drawn surrounding an irregular particle, and if that surface is completely intersected by particle trajectories (collection efficiency much greater than one should suffice), the collection rate will also be independent of shape, for coulombic attraction, which agrees with a result of Levin (see Fuchs [1]).

D. Collection Efficiencies - Central Forces Other than Couloumbic

For the other electrostatic forces, it is not generally true that the divergence is zero. Thus, particle concentrations are not constant along trajectories. Instead, for a central force

$$F = k/r^m \tag{18}$$

in steady-state with no gas flow, the rate of particle passage through a (spherical or cylindrical) surface surrounding the collector would be constant at each distance r (cm):

$$\dot{N} = n(r)w(r)A(r) \tag{19}$$

If the concentration is n_o at $r \geq R_o$, then

$$\frac{n(r)}{n_o} = \left(\frac{r}{R_o}\right)^m \frac{A(R_o)}{A(r)} \quad , \qquad r \leq R_o . \tag{20}$$

For the case in which the velocity v_o is much less than w_s, equation (20) should hold near the collector. Thus for a spherical collector:

$$\frac{n(r)}{n_o} = \left(\frac{r}{R_o}\right)^m \left(\frac{R_o}{r}\right)^2 = \left(\frac{r}{R_o}\right)^{m-2} \tag{21}$$

and for a cylindrical collector

$$\frac{n(r)}{n_o} = \left(\frac{r}{R_o}\right)^{m-1} \tag{22}$$

assuming $w_s >> v_o$. Often collection is put in terms of the free-stream streamline farthest from the flow centerline that is cleaned, a distance R^* such that

$$\dot{N} = n_o v_o \pi R^{*2} \tag{23a}$$

or

$$\dot{N} = n_o v_o 2R^* L_c \tag{23b}$$

for a sphere or a cylinder (of length L_c), respectively. The assumption $w_s >> v_o$ implies that $R^* >> R$, where R is the collector radius, which makes plausible the assumption

$$n(R^*) \doteq n(R_o) = n_o . \tag{24}$$

The concentration at R^* is approximately the free-stream concentration.

For a sphere, this implies

$$\dot{N} = n_o v_o \pi R^{*2} = n(R)w(R)A(R) \tag{25a}$$

$$\doteq n_o \left(\frac{R}{R^*}\right)^{m-2} \left(\frac{Bk}{R^m}\right) (4\pi R^2) \tag{25b}$$

Thus,

$$R^* \doteq (4Bk/v_o)^{1/m} \tag{26}$$

One obtains the collection rate from equation (23a) and the collection efficiency from

$$\eta = (R^*/R)^2 \doteq (4Bk/R^m v_o)^{2/m} \tag{27}$$

which is equivalent to

$$\eta \doteq (4w_s/v_o)^{2/m} = (4K)^{2/m} \tag{28}$$

This is a general expression for strong central force collection efficiency for a sphere, derived without reference to a specific flow or a specific force $(m \neq 0)$.

For a cylinder, a similar analysis yields

$$\eta = (R^*/R) \doteq (\pi Bk/R^m v_o)^{1/m} = (\pi K)^{1/m} \tag{29}$$

a general expression for strong central-force collection efficiency for a cylinder. This approach is believed to be original.

In the other limit, $w_s << v_o$, the collection velocity is much less than the free-stream velocity. Because the collection velocity is low compared with convection (v_o), the concentration near the collector surface should be quite nearly the free-stream concentration:

$$n_s \doteq n_o, \tag{30}$$

in which case the collection rate becomes:

$$\dot{N} = n_o w_s A_s \ . \tag{31}$$

The collection efficiency for a spherical collector will be

$$\eta = 4w_s/v_o = 4K \tag{32}$$

and for a cylindrical collector

$$\eta = \pi w_s/v_o = \pi K \tag{33}$$

Equations (32) and (33) may also be correct for $w_s \sim v_o$ when the gas is sufficiently turbulent to mix the particles well in the vicinity of the collector, because the equations are based on approximating $n(R)$ by n_o. Turbulence might, thus, raise collection.

Intermediate cases $(K = w_s/v_o \sim 1)$ should have efficiencies intermediate between the $K >> 1$ and the $K << 1$ cases studied above, because the concentration at the collector surface should be between n_o and the n(R) calculated for $K \gg 1$.

Approximate efficiency formulas have been obtained here without recourse to trajectory calculations and without assuming particular flow profiles, assumptions having been made about concentration profiles instead. The equations apply to any force such that $F = k/r^m$.

RESULTS: COMPARISON WITH OTHER EFFICIENCY CALCULATIONS

Table 1 is based partly on a similar table by Whitby and Liu (2). The first column has the collector shape. The second indicates which force is being isolated; the subscript Q means the collector is charged and the subscript q means the particle is charged. The third column shows the exact or approximate form of the radial dependence of the central force. The fifth column has the efficiency calculated by the approximations above for K range shown in the fourth column. The last three columns give the efficiencies obtained by others, the flow conditions they assumed, and the reference.

Where there are other determinations with which to compare them, the approximations made here agree rather well (except for F_{oq} at $K \ll 1$, where the central force expression does not remain valid).

It can be noted, in general, that as K increases, the efficiency goes from being linear in K to being sublinear in K, except for coulombic attraction, which means that as steps are taken to raise K, a law of diminishing returns sets in, η increasing less rapidly with K.

DISCUSSION

This simplified approach for the calculation of the single collector electrostatic collection efficiency neglects differences in flow profile, collection by particle interception, and particle inertia. These aspects, and others, are discussed next.

A. Flow Profiles

Kraemer and Johnstone [3] and others [4, 8] have found that for coulombic attraction without interception for inertialess particles, the collection efficiency is independent of whether the flow profile is potential or viscous. Our analysis shows coulombic collection of inertialess particles (even with interception) to be independent of the flow profile whenever the critical particle trajectory strikes the collector directly in the rear (rather than when part of the collector is not intersected by particle trajectories).

For non-coulombic attraction, the results of Kraemer and Johnstone and others indicate some dependence of collection efficiency on flow profile. Results [3] of calculations for collection of uncharged particles on a charged spherical collector were nearly indistinguishable for potential and viscous flows. For charged spherical particles, and an uncharged spherical

Table 1. Efficiency Determination Results

Collector	Force	r^{-m}	K	Collection Efficiency Determinations: This Approximation	Other Efficiency Determinations: Efficiency	Other Efficiency Determinations: Flow	Ref.
Sphere	F_{Qq}	r^{-2}	all	$4K$	$4K$	potential	[3]
	F_{Q0}	r^{-5}	$K>>1$	$(4K)^{2/5}$	$(5.9K)^{2/5}$	potential	[3]
			$K<<1$	$4K$	$\approx 4K$	potential and viscous	[3]
	F_{0q}	$\approx r^{-5}$	$K>>1$	$(4K)^{2/5}$	-	-	
			—		$2.9(K)^{0.35}$	potential	
			—		$1.6(K)^{1/2}$	viscous	[4]
			$K<<1$	$4K$	$1.5(K)^{1/2}$	viscous	[3]
Cylinder	F_{Qq}	r^{-1}	all	πK	πK	potential	[3]
			all	"	πK	all$(r_p/R<<1)$*	[5]
	F_{Q0}	r^{-3}	$K>>1$	$(\pi K)^{1/3}$	$(\frac{3\pi}{4}K)^{1/3}$	potential	[3]
			"	"	$(\frac{3\pi}{2}K)^{1/3}$	all$(r_p/R<<1)$*	[5]
			$K<<1$	πK	πK	"	[5]
	F_{0q}	$\approx r^{-2}$	$K>>1$	$(\pi K)^{1/2}$	$(\frac{4K}{2-\ln Re})^{1/2}$	viscous*	[5]
			"	"	$(6\pi K)^{1/3}$	potential*	[5]
			$K<1$		$1.5(K)^{1/2}$	viscous (experiment)	[6]
			$K<<1$	πK	"	"	[6]
			—		$2.3(K)^{1/2}$	"	[7]

*Natanson's condition was that capture could occur at the rear of the cylinder: $-BF(r_p+R)/v(r_p+R)\geq 1$ at $\theta=0$.

collector, the Kraemer and Johnstone [3] calculations showed that the collection efficiencies were two to three times higher for potential flow than for viscous flow (the streamlines are farther from the collector in viscous flow), for K in the range 10^{-4} to 10^{1}. Other profiles might produce greater differences in efficiency for the same K, especially flows at sufficiently high collector Reynolds number to produce a turbulent wake behind the collector.

B. Particle Interception

Calculations [3] for spherical collectors and spherical particles showed interception to be negligible for $K >> r_p/R$ and dominant for $r_p/R >> K$, for inertialess particles.

C. Particle Inertia and Impaction

Particle inertia can lead to impaction of the particles; the particles may also delay significantly in reaching the electrostatic terminal velocity.

Grover and Beard [9] calculated coulombic collection efficiencies for charged cloud drops and charged particles. They did this by trajectory determinations, using flow fields which were calculated by solving the Navier-Stokes equations, rather than using the viscous or potential flow approximations. (They found that the critical trajectories for inertialess particles did indeed strike the collector directly in the rear). The expression

$$\eta = 4K \tag{34}$$

was found by them to be correct within a factor of two for collector Reynolds numbers less than 100 and for inertial parameters

$$\text{Stk} = v_o\tau/R < 0.2 \tag{35}$$

for the range of particle and droplet charge levels studied. This confirms the restriction on the impaction parameter:

$$\text{Stk} << 1 \ . \tag{36}$$

The same work indicated that as Stk becomes much larger than 1, the collection efficiency became dominated by impaction.

That raises another interesting point. The particle collection rate for the collector with combined impaction and electrostatic attraction will be very approximately the sum of the two collection rates:

$$\dot{N} \doteq n_s w_s A_s + n_o v_o \eta_I(\text{Stk})A_c \tag{37}$$

where η_I (Stk) is the collection efficiency due to impaction alone. The collection efficiency becomes

$$\eta = \frac{n_s w_s A_s}{n_o v_o A_c} + \eta_I(\mathrm{Stk}) \qquad (38)$$

The electrostatic efficiency will decrease as v_o increases, and the impaction efficiency will increase. Although this means there is a minimum collection efficiency at some non-zero v_o, it is misleading. The particle collection rate should increase monotonically as v_o is increased, even when electrostatic attraction dominates impaction. For the same gas volume flow rate and the same total collector area, this would indicate total collection can be increased by making the entire collection zone of the control device have a smaller cross-sectional area perpendicular to the mean flow and a longer dimension parallel to the mean flow. This should not be pursued to extremes, however, because the pressure drop across the collector will also increase as the geometry is changed in this way.

D. Determining Whether Electrostatic Forces are Important

Collection efficiencies are not generally additive, but the effect of two or more collection mechanisms should be greater than the largest of any of them and less than their sum. The removal of particles from the gas means creating a net particle velocity perpendicular to the mean flow sufficient to cause the particles to strike and be held by the collection surface. The total collection surface will be the collector surfaces plus the inner surface of the device's collection zone, that volume that contains the collectors. A first approximation for comparing the magnitudes of the collection mechanisms is to estimate the effective velocity at the surface of a typical collector due to each of the mechanisms; these approximate velocities are as follows:

- impaction: v_o Stk
- interception: $v_o\ r_p/R$
- gravitation: $g\ \tau$
- diffusion: D/R (D is particle diffusivity, cm^2/s)
- electrostatics: w_s

If the estimated electrostatic velocity, w_s is similar to or larger than the largest of the other estimated velocities, then it should be included in the efficiency estimate.

The magnitudes of the particle migration velocity w due to electrostatic forces will clearly depend upon details of particle charge and size and collector charge and size. Figure 2 shows these velocities [13] as a function of particle size under the following assumptions: the particles, if charged, are charged to saturation in an electric field of 10 kV/cm; the collector, if charged, produces a field of 10 kV/cm at its surface and is a 100 μm spherical collection (not grounded). The particles are unit density spheres of the indicated diameter, For the space charge case, the particles were assumed to be charged as above, monodisperse at the size indicated, and having a concentration of $1 \mathrm{g/m}^3$. The migration velocities calculated in this way ranged from $\sim 10^{-3}$ cm/s to $\sim 10^{2}$ cm/s for particle diameters from 0.1 to 3 μm. In decreasing order the velocities were as follows (at 1 μm diameter): charged particle and charged collector (coulomb attraction), charged collector and uncharged particle, charged particle and uncharged collector, space charge repulsion.

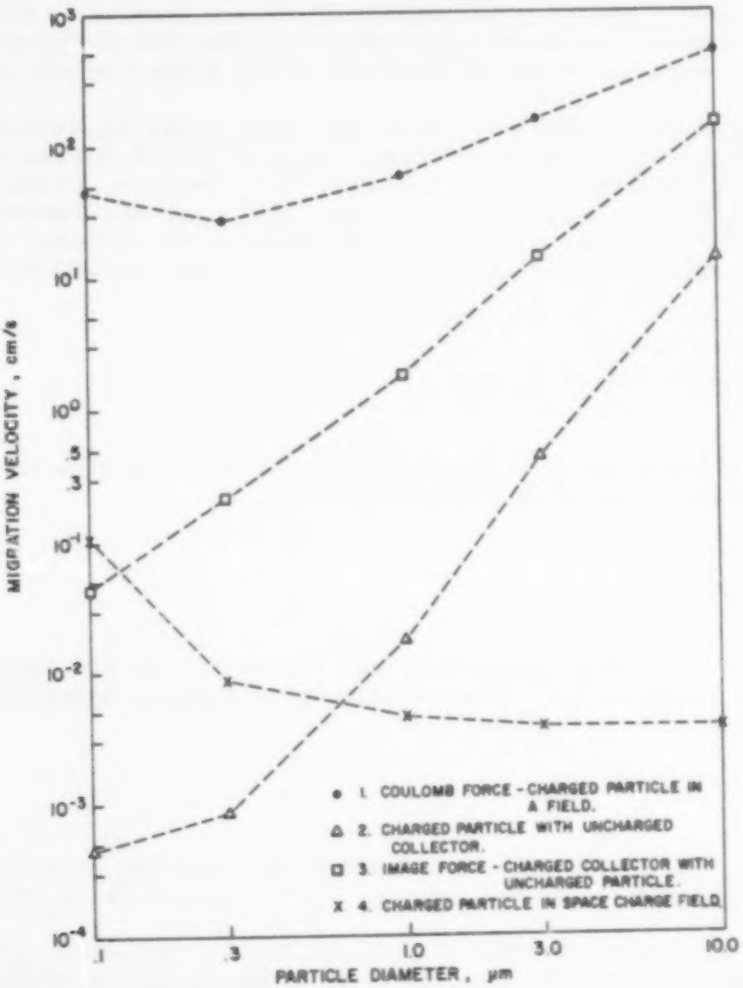

Fig. 2 Migration velocities (particle velocity at collector surface w_s) calculated under conditions indicated in text.

E. Other Fields Superimposed

To a first approximation, the addition of a gravitational field should produce an increase in collection rate due to a gravitational collection rate

$$\dot{N}_g = \bar{n} A_h v_g = \bar{n} A_h g \tag{39}$$

the product of the mean concentration, the horizontal area of the collectors and duct, and the gravitational settling velocity of the particles. The situation is actually more complicated, especially where the flow is either predominantly upward or downward; gravitation can then act as though it changes v_o, allowing the particles more or less time to be in the vicinity of collectors.

An external electric field will not only produce a net migration velocity equal to the product of the particle charge (q_p), the field (E) and the particle mobility (B), but the field will also polarize the collectors, which will in turn modify the field in their vicinity. Zebel [10] gives the following equation for the radial component of field in the vicinity of a cylinder of dielectric constant e_c placed in a field perpendicular to its axis, a field having a strength E_o:

$$E_r = \left(1 + \frac{e_c - 1}{e_c + 1} \frac{R^2}{r^2}\right) E_o \cos\theta \tag{40}$$

The angle θ is measured from the direction of $\vec{E}_o$. The directional nature of the field produced means that the integration

$$\dot{N} = \oint_0^{2\pi} n_s q_p E_r B L_c R \, d\theta \tag{41}$$

is complicated. To a first approximation, the collection efficiency should be between what it would be for a field of strength E_o alone (integrating only over the region where $q_p E_o \cos\theta < 0$)

$$\dot{N}_{E_o} = n_o q_p E_o BLR \tag{42}$$

and twice this, what it would be for a field of strength $2E_o$, the maximum value of E_r, achieved where the collectors are conductors, $e_c \to \infty$.

Nielsen [8] studied the combination of coulombic attraction, gravitation, and external electric field for the collection of inertialess particles on circular cylinders. He found the collection efficiency to be the same "under many conditions for a wide variety of flow profiles, which includes potential, Oseen, and stationary-vortex flows." As one expects from the constancy of particle concentration along trajectories, Nielsen found the deposition density on the cylinder to be uniform for the case of coulombic attraction alone. Combining coulombic and external electrical fields did not necessarily improve collection. Addition of gravitation and an external electric field gave maximum collection when they were perpendicular and minimum when they were aligned.

If collectors are inserted into a region having an electric field, the presence of the collectors will alter that field. If the field is maintained by applying a constant voltage difference, for example across parallel plates, then the field in the gaps amid the collectors will be larger than it would be without the collectors. If the voltage difference V is applied across two parallel plates or grids, the spacing of which is H, then the electric field will be V/H if there are no collectors between the plates. If cylindrical collectors are placed between the grids, with the collector axes parallel to the plane of the grids, then the electric field in the gaps amid the collectors becomes [11]:

$$E = (V/H)/(1-\alpha/4\sqrt{3})$$

in which expression α is the cylindrical collector volume fraction. Thus the field is increased in the gaps as the volume fraction of collectors increases.

If, on the other hand, the external field is due to fixed charges (rather than fixed voltages), the effect becomes the opposite: the more collector volume between the charges creating the field, the less field in the gaps amid the collectors. This shielding is similar to that of a "Faraday cage" and can be quite nearly complete when the collectors are conductive.

F. Cooperative Effects: Many Particles

If the particles are all of like sign, they will repel each other, a phenomenon called "electrostatic scattering" or "space charge repulsion." It can be shown that the force on an individual particle near a conductive collector is just the same as coulombic attraction would be if the collector had charge located at its center equal to the product of the collector volume and the aerosol charge per unit volume. Thus, the efficiencies of collection for spheres and cylinders are 4K and πK, respectively.

If there are only charged particles in a circular duct, and the particles are all of the same sign in charge, then the electric field at the inner surface of the duct is given by the volume integral

$$E = [\oint n\bar{q}_p dV]/2\pi R_d L_d \qquad (43)$$

$$E = \overline{nq}_p R_d/2 \qquad (44)$$

where $\bar{q}_p$ is the mean particle charge and R_d the duct radius. The deposition rate at the inner surface of the duct will be

$$\dot{N} = \bar{n}\overline{q_p BE} A_d/2 = n\bar{q}_p \overline{Bnq}_p 2\pi R_d L_d/2 \qquad (45)$$

or

$$\dot{N} \doteq n^2 \bar{q}_p^2 \bar{B} 2\pi R_d^2 L_d/2 \tag{46}$$

The presence of particles of opposite charge, collectors of opposite charge, or even collectors which are uncharged will all serve to lessen collection due to space charge repulsion at the duct inner surface. Correct calculation of the space charge field within a filter or scrubber requires taking into account in detail the geometry, particle charge concentration, collector volume fraction, collector charge, and collector dielectric constant. This effect may be important when, for example, non-conductive fibers collect unipolarly charged aerosol particles.

If the particles are of equally positive and negative charge, then they will coagulate more readily, but the net field at the inner surface of the duct will have a zero contribution from particle charge.

G. Cooperative Effects: Many Collectors

In general, if there are many collectors (droplets, fibers, etc.), the collection efficiency of the assembly is

$$E \doteq 1-\exp(\ WA/Q) \tag{47}$$

where

$$WA = \sum_i (w_s A_s)_i$$

that is, WA is the sum of the products of the collection areas and the migration velocities. Q is the volume flow rate through the collector. The presence of other collectors will change somewhat the gas flow in the vicinity of a given collector, so that its collection efficiency will be somewhat different from that calculated assuming potential or viscous flow. As we have seen, the collection efficiencies for electrostatic effects have not been strongly influenced by details of the flow profile. If the collectors are charged to the same sign, this will enhance the space charge repulsion effect. If they are of mixed charge, that will decrease space charge but enhance collection on the collectors, because the electrical fields will be somewhat greater than those calculated by considering the isolated collector.

H. Cooperative Effects: Loss of Collectors

In filters, the collection surfaces are held fixed with respect to the gas stream to be cleaned. In scrubbers, both collectors and particles move with the gas. Melcher and Sachar [12] analyzed many different charged particle-charged droplet configurations. The rates of change of concentration of particles and collecting droplets were put in terms of characteristic times (times for approximately 1/e decrease in concentration).

For particles,

$$t_a = 1/4\pi\, q^2 Bn \tag{49}$$

is appropriate for particle neutralization due to agglomeration in a bi-polar aerosol or for particle concentration decrease, due to electrostatic scattering, for a unipolar aerosol. For droplet neutralization (if bipolar) or scattering (if unipolar) a similar time can be calculated:

$$t_R = 1/4\pi\, Q^2 B_d n_d \tag{50}$$

where the subscript d indicates these are the mobility and number concentration of the droplets. In order for the droplet charge or concentration not to be greatly diminished before collection is completed, t_R should be large in comparison to the time characteristic of coulombic collection

$$t_c = 1/4\pi\, qQBN_d. \tag{51}$$

Similarly, if the charge on the particles is bi-polar, then for nearly complete collection, t_a must be greater than t_c, which in turn must be much greater than the residence time of the particles in the collection zone.

CONCLUSION

It has been shown for inertialess particles undergoing collection due to central forces that the collection efficiency for a single spherical collector is (neglecting interception):

$$\eta \doteq (4K)^{2/m} \tag{52}$$

for K > > 1, where K is the ratio of the particle terminal velocity at the collector surface, due to a force which varies radially as $1/r^m$, and v_o is the free-stream velocity. The corresponding equation for a cylindrical collector is

$$\eta \doteq (\pi K)^{1/m} \tag{53}$$

When K < < 1, the approximate collection efficiencies are

$$\eta \doteq 4K \tag{54}$$

for a spherical collector and

$$\eta \doteq \pi K \tag{55}$$

for a cylindrical collector. These approximations should hold for any flow profile and any central force of the form $F = k/r^m$.

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ACKNOWLEDGEMENT

Support by the National Institutes of Health through NIH Grant 5-S07-RR-05446-15 is acknowledged.

WRITTEN DISCUSSION

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The particle flux method used by Dr. Cooper, which was developed independently by Chalmers [14] and Dukhin and Deryagin [15], and the stream function method for determining particle trajectories [4, 8, 16, 17], which was first used by Whipple and Chalmers [18], are two powerful techniques for analyzing particle collection behavior when particle inertia can be ignored. Both methods can only be applied, however, to solenoidal particle fields resulting from incompressible flows and solenoidal forces. Such forces include the gravitational force and coulombic and external electric field forces on a charged particle. These electrical forces are given by $Q_p\underset{\sim}{E}$, where Q_p is the particle charge and $\underset{\sim}{E}$ is the electric field intensity present in the absence of the charged particle and, from Maxwell's equations, $\underset{\sim}{E}$ is solenoidal. If the flow, force, and hence particle fields are solenoidal, then for two-dimensional systems stream functions for the fields exist. The stream function for particle trajectories, which provides a complete description of particle motion, is then given by superposition of the flow and force stream functions.

In the particle flux method, the solenoidal property of the particle field makes it possible to calculate the particle flux at the collector surface without having to solve for the particle concentration profile in the flow, because particle concentrations are constant along the particle trajectories. This gives limited but useful information about the particle motion, namely, the collection efficiency. The most useful application of this method is in obtaining results for three-dimensional systems, for which the stream function method can not be used. Whenever both attractive and repulsive forces are present at the surface of the collector (such as for a combination of coulombic and external electric field forces), however, the method must be applied with caution to avoid erroneous results. From just a knowledge of the force distribution at the collector, it is not possible to distinguish between the flux of particles coming from upstream and the flux of particles on closed trajectories originating elsewhere on the collector. This determination requires a knowledge of the flow field and the types of limiting trajectories that occur. Although the method was originally proposed for point particles, it can also be applied to collection with interception, using the force distribution at the surface surrounding the collector that is defined by the centers of particles at capture [17].

The results found by Dr. Cooper for the coulombic force case, given by collection efficiencies of $-4K_{Qq}$ for spheres and $-\pi K_{Qq}$ for cylinders, where K_{Qq} is the coulombic force parameter (negative for attraction), are remarkable in their generality. Furthermore, the result for spheres has been verified experimentally [3] for measurements over a wide range of parameter values ($10^{-3} < -K_{Qq} < 10$). That these results hold for all steady incompressible flows was shown by Nielsen and Hill [4,16] for spheres and Natanson [5] and Nielsen [8] for circular cylinders. That both results are insensitive to both collector shape and flow profile was shown by Levin [18] and Nielsen [17]. In addition, Nielsen also noted that the rate of particle collection on fibers should be unaffected by accumulation of previously captured particles (provided net fiber charge is not significantly affected). Coulombic force influence on particle dendrite formation was also examined.

Using both the stream function and particle flux methods, Nielsen [8, 17] has obtained particle trajectories and collection efficiencies for the external electric field force on a charged particle for cases in which the force acts alone or in combination with the coulombic force and gravitation. Collector geometries used are spheres, prolate and oblate spheroids, disks, circular and elliptical cylinders, ribbons, and collectors of arbitrary shape as well. Flow fields considered are potential, Stokes, Oseen, stationary-vortex, cellular, and arbitrary steady incompressible flows. Results were found for arbitrary external electric field, flow and collector orientations and include the effects of particle interception. In general, the results show that for any given collector shape and given flow and field orientations, for any combination of the forces the collection efficiency is the same for all steady incompressible flows when interception effects are negligible. In cases in which the coulombic force is sufficiently dominant, the collection efficiency is given by $-4K_{Qq}/V_o$ for sphere-type collectors and $-\pi K_{Qq}/V_o$ for cylinder-type collectors independent of the collector shape, of the flow field, and of the orientations of the collector and external field. $V_o = (1 + 2K_{ex}\cos\theta + K_{ex}^2)^{1/2}$ is the ratio of the particle approach speed far upstream to the free-stream velocity, where K_{ex} is the external electric field force parameter (see Table 3) and θ is the angle between the flow and external field. For the external electric field force alone, the efficiency is given by $\eta_{max}K_{ex}/V_o$, where η_{max}, the collection efficiency for infinite field strength, is a function only of collector geometry, dielectric constant, and orientation of the collector with respect to the external field.

The problem of calculating collection efficiencies for nonsolenoidal electrical image forces is much more difficult than for the coulombic force case since the particle concentration profile is not uniform and changes with flow field. When analytical descriptions of a flow field are available it is a simple task to obtain efficiencies via numerical integration of particle trajectories, but the simple flow fields normally used are not very representative of real flows and, hence, the results are often less than satisfying. Dr. Cooper's approximate analyses represent the first attempt to use the particle flux method to obtain results that reflect an "average" influence of flow field. Indeed, if such results were available for the whole range of image force strengths they would be very useful.

In Table 1, Dr. Cooper has assembled results of previous studies for comparison with his high and low efficiency asymptotic results. Unfortunately, Table 1 contains several inconsistencies, misinterpretations, and omissions of pertinent results. The apparent difference in results obtained by Kraemer and Johnstone [3] and Natanson [5] for force F_{Q0} for cylinders comes from parameter definitions that differ by a factor of two. For force F_{0q} for spheres, Nielsen [4] and Kraemer and Johnstone obtained identical results for the viscous flow case and in their analyses used a force expression that goes to $2r^{-5}$ for large r; hence their parameter definitions differ from Cooper's by a factor of two. Natanson's [5] results for force F_{0q} for cylinders were derived for the low and not high efficiency case. In order to avoid confusion for the reader and to include additional results, a revised version of Table 1 is given in Table 2, with results reported on the basis of the electrical force parameter definitions appearing in Table 3, which have been used by Kraemer and Johnstone and Nielsen.

Compared to the coulombic force, the electrical image force (F_{Q0}) between a neutral particle and a charged collector is a weak, short range force. When both forces occur, the image force can be neglected (for spherical collectors) if the particle-to-collector charge ratio is much greater than the volume ratio. However, for droplet collectors charged to near the Rayleigh limit by electrical atomization the image force can be important, but efficiencies of micron sized particles are much less than needed for high efficiency results to be valid. As indicated in Table 2, a number of investigators have obtained results for the charged-collector image force case. In addition, for intermediate parameter values for spheres, results which do not follow a power law are available for potential and Stokes flows [3, 2, 16]. For the high efficiency results ($K >> 1$), the limiting trajectory lies far from the collector, where flow fields all degenerate to uniform flow. Hence these results are independent of flow field and, similarly, independent of collector geometry. The low efficiencyresults ($K << 1$), which agree with those of Dr. Cooper, are said to be valid for all flows, but this results from ignoring radial variations of the force near the collector and hence making the particle field artifically solenoidal. In the exact case results should vary with velocity gradient at the surface; that Kraemer and Johnstone's numerical results for potential and Stokes flows do not merge even at very low parameter values supports this.

The electrical image force (F_{0q}) between a charged particle and a neutral collector has the unique property of becoming infinite at the collector surface, although it drops off very rapidly with distance. It is negligible compared to coulombic interaction (for spheres) when particle charge is much less than collector charge. For micron sized particles collected on droplets, parameter values are typically much less than one. However, the force will always exert some influence on trajectories passing very near the collector surface. The high efficiency limit results are independent of flow field,

Table 2. Comparison of Collection Efficiency Results

Collector	Force	r Dependence*	Parameter Range	Cooper's Results	Other Studies	
					Efficiency	Flow and Reference
Sphere	F_{0q}	r^{-2}	all	$-4K$	$-4K$	incompressible [16, 18]
	F_{Q0}	r^{-5}	$K >> 1$	$(4K)^{2/5}$	$(\frac{15\pi}{8}K)^{2/5}$	uniform [3, 19, 20]
			$K << 1$	$4K$	$4K$	all [21]
	F_{0q}	$\frac{r}{(r^2-1)^2} - \frac{1}{r^3}$	$K >> 1$	$(8K)^{2/5}$	$(\frac{15\pi}{4}K)^{2/5}$	uniform [19]
		$= 2r^{-5}$	$.002 \leq K \leq .1$	-	$(20.2K)^{.353}$	potential [3, 4]
		for $r >> 1$	$.002 \leq K \leq .1$	-	$(2.50K)^{1/2}$	Stokes [3, 4]
			$K << 1$	$8K$	$(\frac{72}{5}K)^{1/3}$	potential [21]
			"		$(2K)^{1/2}$	Stokes [19]
Cylinder	F_{0q}	r^{-1}	all	$-\pi K$	$-\pi K$	incompressible [5, 17, 18]
	F_{Q0}	r^{-3}	$K >> 1$	$(\pi K)^{1/3}$	$(\frac{3\pi}{2}K)^{1/3}$	uniform [3, 5]
			$.01 < K < 1$	-	$(.79K)^{3/4}$	Kuwabara [22]
			$K << 1$	πK	πK	all [5]
	F_{0q}	$(r-1)^{-2}$	$K >> 1$	$(\pi K)^{1/2}$	$(4K)^{1/2}$	uniform**
		$= r^{-2}$	-	-	$(5.29K)^{1/2}$	experimental [7]
		for $r >> 1$	$K < .03$	-	$(2.25K)^{1/2}$	experimental [6]
			$.001 < K < .1$	-	$(2.25K)^{1/2}$	Kuwabara [22]
			$K << 1$	πK	$(\frac{4K}{2-\ln Re})^{1/2}$	viscous [5]
			"		$(6\pi K)^{1/3}$	potential [5]

* r is made dimensionless using collector radius. ** Derived by Nielsen via Natanson's method.

Table 3. Electrical Force Parameter Definitions (K)

	SPHERE	CYLINDER
Coulombic Force (F_{Qq})*	$\frac{CQ_cQ_p}{24\pi^2\epsilon_0 R_p R_c^2 \mu U_0}$	$\frac{C\rho_c Q_p}{12\pi^2\epsilon_0 R_p R_c \mu U_0}$
Charged-Collector Image Force (F_{Q0})**	$\left(\frac{\epsilon_p - \epsilon_0}{\epsilon_p + 2\epsilon_0}\right)\frac{CQ_c^2 R_p^2}{12\pi^2\epsilon_0 R_c^5 \mu U_0}$	$\left(\frac{\epsilon_p - \epsilon_0}{\epsilon_p + 2\epsilon_0}\right)\frac{C\rho_c^2 R_p^2}{6\pi^2\epsilon_0 R_c^3 \mu U_0}$
Charged-Particle Image Force (F_{0q})	$\left(\frac{\epsilon_c - \epsilon_0}{\epsilon_c + 2\epsilon_0}\right)\frac{CQ_p^2}{24\pi^2\epsilon_0 R_p R_c^2 \mu U_0}$	$\left(\frac{\epsilon_c - \epsilon_0}{\epsilon_c + \epsilon_0}\right)\frac{CQ_p^2}{96\pi^2\epsilon_0 R_p R_c^2 \mu U_0}$

External Electric Field Force
(all collector geometries)

$$\frac{CQ_pE_0}{6\pi\mu R_pU_0}$$

Q_c	collector charge	R_c	collector radius
Q_p	particle charge	R_p	particle radius
ρ_c	collector charge per unit length	U_0	free stream velocity
ϵ_0	fluid dielectric constant	μ	viscosity
ϵ_p	particle dielectric constant	E_0	uniform external electric field intensity
ϵ_c	collector dielectric constant		
C	Cunningham-Millikan slip correction factor		

* For collectors of different geometry R_c is an arbitrary collector reference length (an equivalent radius, for instance) upon which the definition of collection efficiency depends.

** For cylinders multiply by 2 to get Kraemer and Johnstone's definition.

but not necessarily independent of collector shape. As noted by Dr. Cooper, it is not valid to use the large distance force approximation near the collector. The literature low efficiency results reflect the actual force near the surface but, as Davies [23] has pointed out, approximations made in their derivations could produce serious limitations on their usefulness. Nevertheless, for cylinders the one half power law for viscous flow, which appears valid over the entire range of parameter values, has received experimental [6,7] and numerical [22] confirmation. For spheres the results for Stokes flow show little deviation from a one half power law [3,4].

When particle inertia is present, second-order effects can create very complex particle motions, and analytical solutions are generally not available. For spherical collectors, however, collection efficiencies for the combined influences of particle inertia and electrical forces have been obtained by Nielsen and Hill [4, 23] through numerical integration of particle trajectories using potential and Stokes flows. The electrical forces considered are the coulombic force, both electrical image forces, and external electric field forces, with the external field parallel to the flow. Results for perpendicular fields and for prolate spheroidal collectors are given by Mukherjee and Hill [24]. These results show that characterizations made on the basis of inertialess analyses (such as flow independence of coulombic force collection) can not be extended into the inertial realm. Although maximum collection is favored by large particle inertia when electrical forces are absent, collection with electrical forces often reaches a maximum for negligible particle inertia. In general, the steeper the velocity gradient at the collector, the greater the effect inertia has. The statement by Dr. Cooper on the additivity of purely inertial and purely electrical collection efficiencies (Equation 38) simply is not borne out by these studies even in approximate form.

Although analyses of particle collection on single, isolated droplets and fibers are very similar, the dynamic characteristics of droplets in wet scrubbers and fibers in filters are very different. In wet scrubbers, droplets follow the bulk flow and only affect a localized region. Shadow or distortion effects caused by the particle field interacting successively with several droplets is minimal, and isolated collector results may be a good approximation. Whereas, in filters fibers continually sweep the bulk flow, and upstream fibers can cast particle shadows on downstream fibers or otherwise distort the particle field such that simultaneous analysis of multiple fibers is really required. This is a difficult task, however, for which new insights into particle collection behavior, such as found in Dr. Cooper's approximate analyses, may be required.

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OPEN DISCUSSION

Friedlander: We have heard a criticism with high gradient magnetic separation theory. It may be interesting to point out that there is no central force in the HGMS theory and that complicates matters tremendously. This is very clearly demonstrated. Calculations can be made rather readily in the case of the electric field. In the case of the magnetic field, there are always dipolar interactions which never have central forces. That is what makes matters so much worse. On the other hand, in the situations so far reported, the collection agent is usually stationary. That will change as one goes to gaseous streams.

Cooper: A quick comment. One of the ways you can pursue it, at least to an approximate level is to find the directional force and the area of the particle collectors which indeed collect. You then have some idea of what the velocity is on the surface of that area. You have to make an assumption about the concentration profile.

Löffler: I agree completely with your remarks and that you expect from theory to get some ideas on the most important inputs and so on. That is the same point of our theoretical work.

There is one point on your assumptions and simplifications that may cause a problem. I agree as long as particles are below .1 μm but as I have shown earlier they may range between .1 and 1 μm. You should think of the supposition you used to simplify your equation. Another thing is that all these comments also refer to the interception effect, and I must insist that interception is not an impact force mechanism. Interception is only the contact between particle and collector. You might speak of interception if the mass densities of particles in the fluid are equivalent. Then, you have some effect, but it is not an impact force mechanism. Sometimes it is important, especially in your case where the particle is collected in the rear.

Melcher: On the question of impaction and interception, the paper by Brogan & Beer doing for the spherical case what Professor Loffler has done for the cylindrical case, may be of interest. Essentially you want to take collection efficiency as a function of velocity. For fairly weak electrical forces you are going to have decreasing collection efficiency. Then you will have the typical impaction curve and you will get a combined thing. It will come off about 1 or a little higher if you want to take interception into account. If you have a very strong electrical force, you will eventually have the collection efficiency asymptotically approach the impaction collection efficiency. It is a little misleading, however, because if you look at the collection rate, it turns out to be equal to the concentration on the surface times the velocity at the surface times collecting area of the surface plust the impaction contribution.

The rate of collection will remain almost independent of velocity. There is a Reynolds number correction or if we get too high, there is a problem of collecting particles with great inertia. The last thing we have is the freestream concentration which will not change, The velocity will go up. As it turns out, with the same collectors and volume flow, you want to arrange the geometry that as long as you can handle the pressure drop, the velocity is high. That suggests that going from a stubby collection area to one that is small and long, both having the same volume flow. Indeed then you can get the impaction.

The other point is that, after a while, the particles get so massive that you just cannot pull them in with electrical forces. Impaction does work well.

Liu: I understand your assumption, the divergence of the force will be equal to zero, which leads to the conclusion that the concentration of particles along the trajectory remains constant.

Cooper: That is for the first part of the paper where I deal with special cases of Coulombic force, gravitational force and uniform electric field. We stopped that development at that point, then took away that assumption and dealt with central forces in general. The rest of the paper deals with a generalized central force.

Liu: The first part of your paper would also be applicable to the magnetic situation, because you have a uniform particle concentration along the trajectory even for the case of dipole magnetic particles. Because the divergence of the field will be zero.

ELECTROSTATIC FILTRATION AND THE APITRON/ DESIGN AND FIELD PERFORMANCE

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and

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ABSTRACT

The electrostatic augmentation of fabric filtration has been examined for many years. Most often these studies have been directed toward the effects of electrostatics on the efficiency of particle collection, and it has generally been shown that efficiency increases with the addition of electrostatics. It is shown in this paper that electrostatic augmentation also gives rise to substantial increases in filtration rate per unit fabric area. An industrial fabric filter utilizing electrostatics is described, and its performance is demonstrated on several typical applications.

INTRODUCTION

The effects of electrostatics on the fabric filtration process have been investigated and discussed for decades. In 1930, N. J. Hansen discovered that the filtering efficiency of a wool filter pad was much improved when powdered with ground resin because the resin possessed a substantial static charge (1). This discovery was used to improve the particulate collection efficiency of military gas masks. Since that time, many studies have been made concerning the effects of charged particles, charged fibers, externally generated electrostatic fields, or combinations of these factors.

Frederick (2) studied the interactions between naturally charged dusts and fabrics. Lundgren and Whitby (3) investigated the effect of particle charge on the filtration efficiency of uncharged fibers. Walkenhorst and Zebel (4) applied an electrostatic field across a fabric filter. Inculet and Castle (5) describe a device in which particles are charged and a field is applied across the filter. The work of Lundgren and Whitby (3), shown in Figure 1, illustrates the magnitude of the influence of electrostatics on fabric filter collection efficiency.

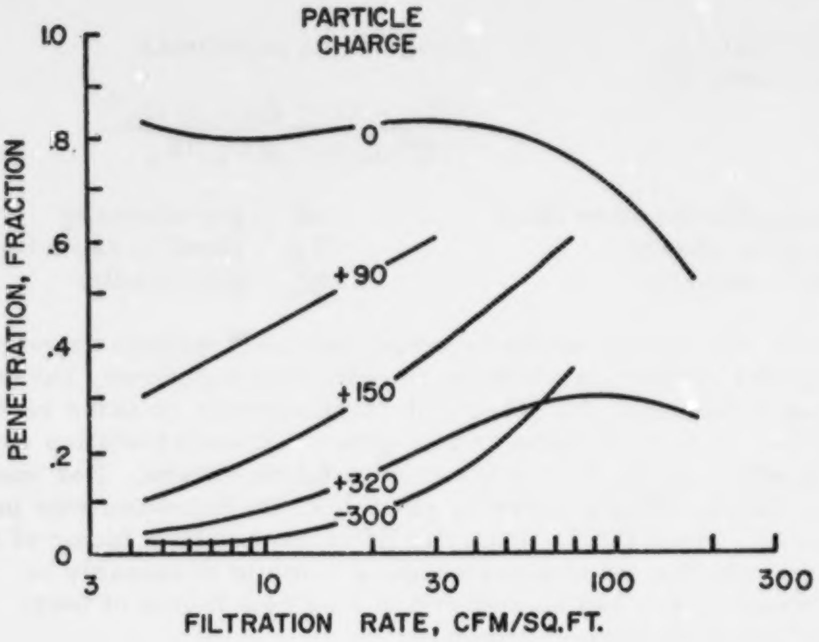

Figure (1) - Particulate penetration through a filter as a function of particle charge and filtration rate (from Ref. 3)

Mathematical models for the efficiency of fabric filters have been presented by several investigators. Natanson (6) gives equations for the collision efficiency between a charged particle and a dielectric fiber. Zebel adds an external electrostatic field to the same problem and presents solutions to the governing equations of particle motion. Ariman and Tang (8) extend Zebel's results by including the effects of neighboring fibers.

Virtually all of these investigations, both experimental and theoretical, indicate that a substantial increase in collection efficiency can be realized when particles are charged or fields applied. This increased collection efficiency is the result of induced polarity of the fibers or particles, which effectively increases the collision cross section between a fiber and an approaching particle. Experimentally, this effect appears to obey the equation (1, 3)

$$\eta = \text{single fiber collision efficiency} = \alpha (K)^{\frac{1}{2}}$$

$\alpha = 2.3$ (ref. 1)
 1.5 (ref. 3)

K = image force parameter

$$= \frac{(\varepsilon_c - 1)/(\varepsilon_c + 1)\, Q_p^2}{12\pi^2 \varepsilon_o D_f^2 \mu D_p v_o}$$

ε_c = dielectric constant fiber
Q_p = particle charge
D_f = fiber diameter
μ = gas viscosity
D_p = particle diameter
v_o = gas velocity

It is obvious from the results so far reported that electrostatic augmentation of fabric filters yields improvement in collection efficiency, but very little information concerning the effects of electrostatics on filter resistance is available. It will be shown in this paper that electrostatics can have a profound effect on the flow resistance of fabric filters. For example, it will be shown that by simply charging particles, the filtration rate per unit fabric area of a given fabric filter can be increased by a factor of four. This implies that with the aid of electrostatics it might reasonably be expected that fabric filters can be reduced in size by a factor of four, while achieving better collection efficiency.

BASIC STUDIES

The results of recent bench scale tests can best be used to illustrate the variation of filter resistance with particle charge. At constant face velocity, suspended dusts were filtered by initially clean fabrics. The dusts were charged prior to deposition by means of passage through a corona discharge. Degree of charge was controlled by the magnitude of the field strength, and pressure drop was recorded after a specified weight of dust had been deposited.

Filter resistance, or pressure drop, decreases as the degree of particle charging increases, as is shown in Figure 2.

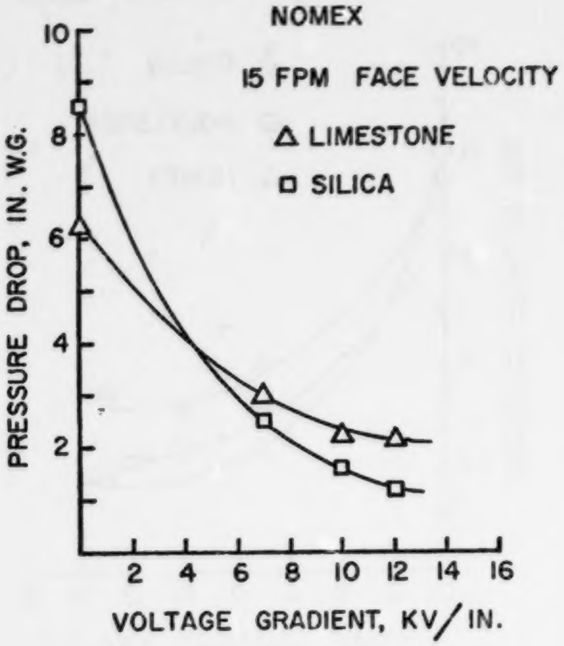

Figure (2) - Pressure drop vs. charging voltage gradient for the deposition of limestone and silica dusts on Nomex felt

Each point of this figure represents the pressure drop resulting from the deposition of 25 grams of dust at a 15 foot per minute face velocity. In particular, for the silica dust, it can be seen that filter resistance is decreased by a factor of 8. Although there are quantitative differences in the magnitude of the effect of charging dusts prior to filtration, the effect is qualitatively identical for all dusts tested.

Fabric type exerts some influence on the electrostatic filtration process, but the relative differences among fabrics are small, as is shown in Figure 3. Charging dust particles always results in a substantial reduction of pressure drop, regardless of fabric type.

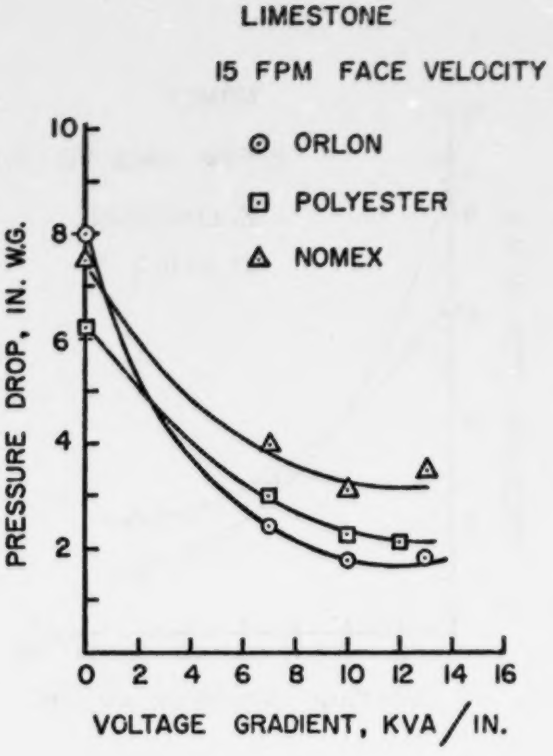

Figure (3) - Pressure drop vs. charging voltage gradient for the deposition of limestone dust on three felted fabrics

The variation of filter resistance with face velocity is shown in Figure 4. Each point of this figure represents the pressure drop resulting from the deposition of 25 grams of dust at the face velocity given. Pressure drop increases with face velocity as expected, and it can be seen that particle charging results in a two-fold increase in face velocity at an equivalent pressure drop.

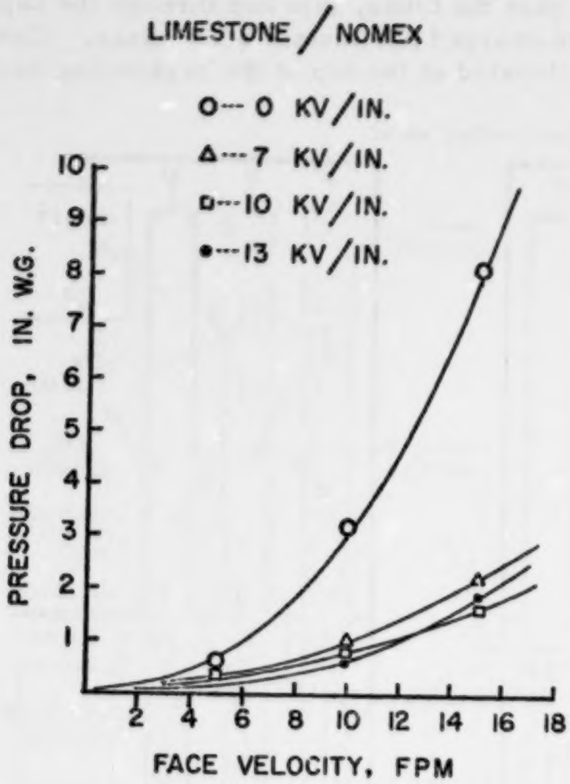

Figure (4) - Pressure drop vs. deposition face velocity for limestone dust on Nomex felt as a function of charging voltage gradient

FILTER DESIGN

In order to apply electrostatics to the fabric filtration process, efforts were initiated in 1971 for the design of a practical device. Various configurations were tried, and the geometry of Figure 5 was ultimately chosen as most acceptable. This device is designated the electrostatic fabric filter or "Apitron". During normal operation, incoming airflow enters the precipitator section in the area below the walkway. It is deflected downward and distributed among the tubes by means of an inclined, expanded metal grating. The flow then passes upward, through

the tubes, where the particulate is charged and where most is deposited. Flow continues upward past the tubes, into and through the bags, where the final filtration of the charged particulate takes place. Clean air exits the unit at the exhaust, located at the top of the baghousing section.

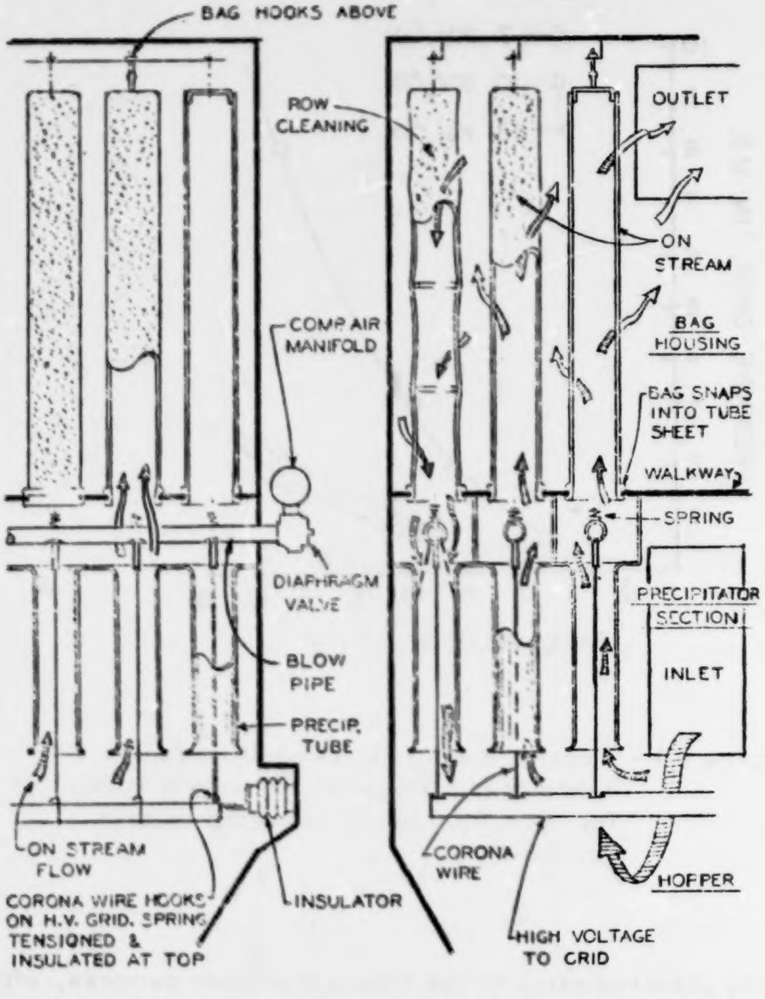
BAG HOOKS ABOVE
ROW CLEANING
OUTLET
ON STREAM
COMP AIR MANIFOLD
BAG HOUSING
BAG SNAPS INTO TUBE SHEET
WALKWAY
SPRING
DIAPHRAGM VALVE
PRECIPITATOR SECTION
BLOW PIPE
INLET
PRECIP. TUBE
ON STREAM FLOW
CORONA WIRE
HOPPER
CORONA WIRE HOOKS ON H.V. GRID, SPRING TENSIONED & INSULATED AT TOP
INSULATOR
HIGH VOLTAGE TO GRID

Figure (5) - The Apitron industrial scale electrostatically augmented fabric filter

Periodically, the deposit of particulate is cleaned from the tube and fabric. Six bags and tubes are cleaned of deposited dust at one time. This cleaning is initiated when an electrical pulse from the control box opens one

of the twelve diaphragm valves for 1/10 second. The blow pipe connected to the valve is then pressurized and compressed air jets downward from six nozzles, each directly above a tube and concentric to a corona wire. The jet of air flowing downward through the tube entrains and mixes with a secondary airflow, and the tube is swept clean of deposited dust by the mixture of high velocity air. The secondary airflow, passing from the outside to the inside of the bag, snaps the bag inward and dislodges dust deposit from it. Typically, a single bag and tube would be cleaned in this manner once every five minutes.

PERFORMANCE

When no high voltage is applied, the fabric filter behaves as a conventional continuous cleaning, pulse type baghouse. As voltage is applied, filter resistance decreases as shown in Figure 6.

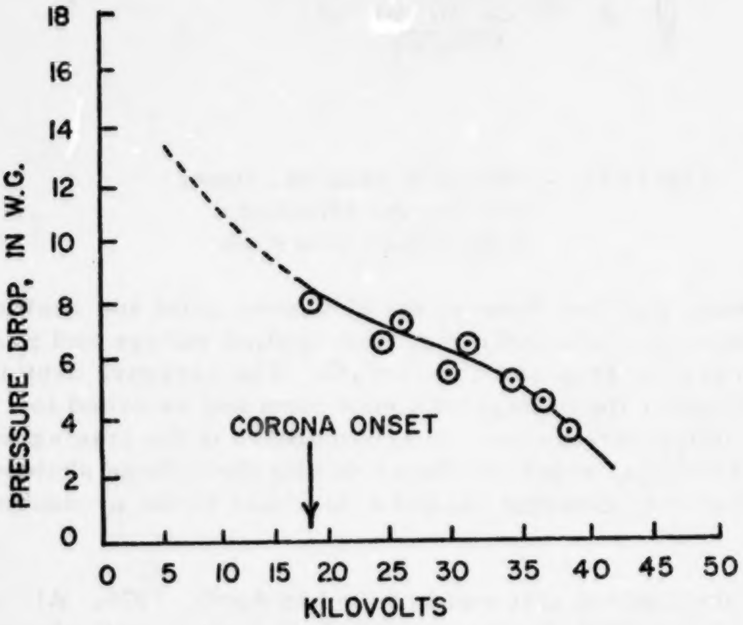

Figure (6) - Pressure drop vs. charging voltage for the Apitron collecting silica dust

This decrease in resistance continues as voltage is increased because particles become more highly charged and because more particulate is deposited on the metal tube. This behavior is similar to that shown in Figure 2 for the bench scale tests.

Figure 7 shows a recorder tracing of pressure drop during the collection of silica at a 15 CFM/Sq. Ft. filtration rate.

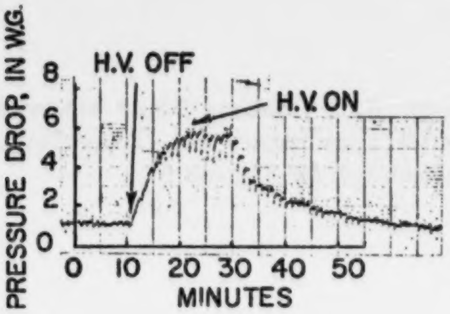

Figure (7) - Pressure drop vs. time, showing the effect of a high voltage shut down

The applied voltage was shut down at the 11 minute point and restored at 23 minutes. Under normal conditions, with applied voltage and particle charging, the pressure drop was 1 inch W.G. The pressure drop increased to 6 inches W.G. after the voltage was shut down and returned to 1 inch following high voltage restoration. Discontinuities in the tracing were caused by bag cleanings, which continued during the voltage shutdown, and it can be seen that each cleaning caused a decrease in the pressure drop as expected.

Field testing of the Apitron unit was initiated in April, 1976. All tests are carried out on a 1000 CFM, trailer-mounted pilot plant unit, shown in Figure 8. A slipstream of process gas is withdrawn from exhaust ductwork and is then filtered by the pilot plant unit. Generally, this operation is carried out in parallel with existing emission control systems, often baghouses, and comparative performance can be easily defined. Each field test is run for two weeks or longer, and stable operation is achieved before data is taken. Occasionally, dilution air or water spray is added to the pilot plant inlet in order to reduce temperature or to affect some degree of gas conditioning.

APITRON
AMERICAN
PRECISION
INDUSTRIES, INC.

Figure (8) - Apitron mobile pilot plant at asphalt batching test location

The results are shown in Table 1, and Figures 9 and 10 are examples of typical data. The performance of parallel operating baghouses is shown, and it can be seen that air cloth ratios from four to seven times higher than that of conventional equipment can be achieved. In addition, collection efficiency is generally improved, resulting in a 1/2 to 1/10 reduction of particulate emissions on a weight basis.

TABLE 1

Application	Air Cloth Ratio * CFM / Sq. Ft.	Inlet Loading Grain / ACF	Outlet Loading Grain / ACF	Particulate MMD Microns
Asphalt Plant Rock Dryer	12.5	32	.002	> 40.0
Gypsum Rock Dryer	13.5	10	.001	7.0
Phosphate Conveyors	13.0	7	.0008	12.0
Electric Steel Furnace	8.0	.13	.0008	2.5
Asbestos Wallboard Finishing	15.0	.32	.0004	Fibrous ~3.0
Cement Kiln	8.0	3.5	.004	2.8

* Measured at a 3-inch W. G. Pressure Drop

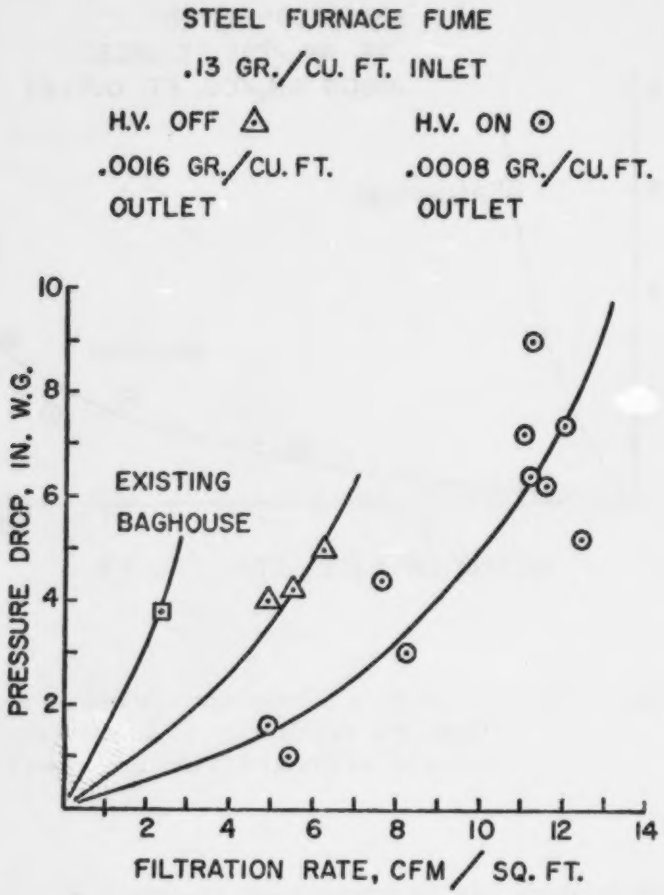

Figure (9) - Results of pilot plant testing on fume from an electric steel furnace, showing Apitron performance with and without electrostatic augmentation

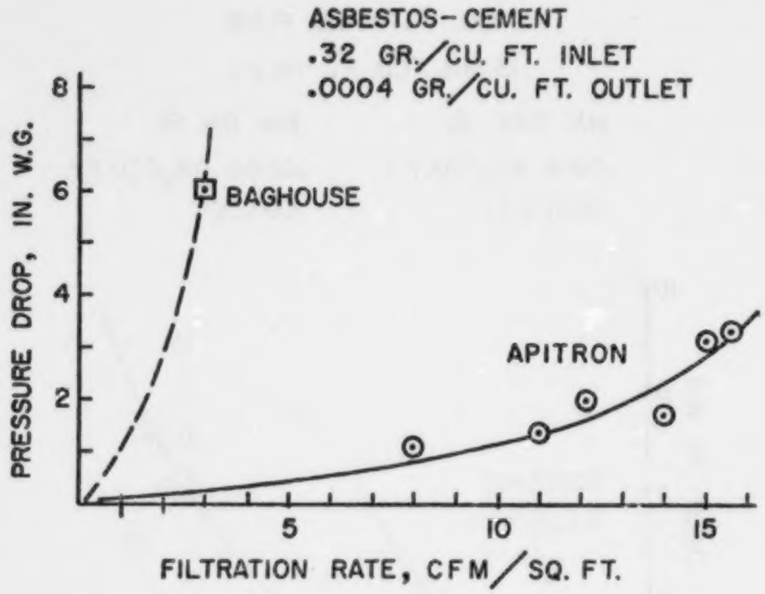

Figure (10) - Results of pilot plant testing on dust from the ventilation of an asbestos cement wallboard trimming operation

CONCLUSIONS

Two basic conclusions can be drawn from the given results: high collection efficiencies can be expected from the Apitron, and high air cloth ratios can generally be achieved. As previously discussed, improved collection efficiency results from the electrostatic augmentation of the filtration process because of polarization effects between charged particles and fibers.

A more surprising result is that of the demonstrated increases in filtration capacity per square foot of fabric when charged particulate is collected. This occurs because the particulate deposits on the fabric in a more open, loosely packed structure, and this difference in structure can be observed under magnification. The structure is more permeable and, hence, allows greater filtration rates.

For a given flow rate of gas, the filtration capacity, or air cloth ratio, directly determines the overall dimensions of a fabric filter because it specifies the area of fabric needed. The method and frequency of removal of the deposited particulate from the fabric can effect air cloth ratio requirements. For example, a fabric filter utilizing a shake cleaning mechanism might require an air cloth ratio of 2, while a jet pulse fabric filter would yield adequate performance on the same application at an air cloth ratio of 5. With the addition of electrostatics, the air cloth ratio could be increased to 14. Typically, an air cloth ratio is selected such that the pressure loss of the fabric filter will fall between three and six inches of water. Figure 11 shows the air cloth ratio capabilities of three types of fabric filters and the resulting floor space requirements.

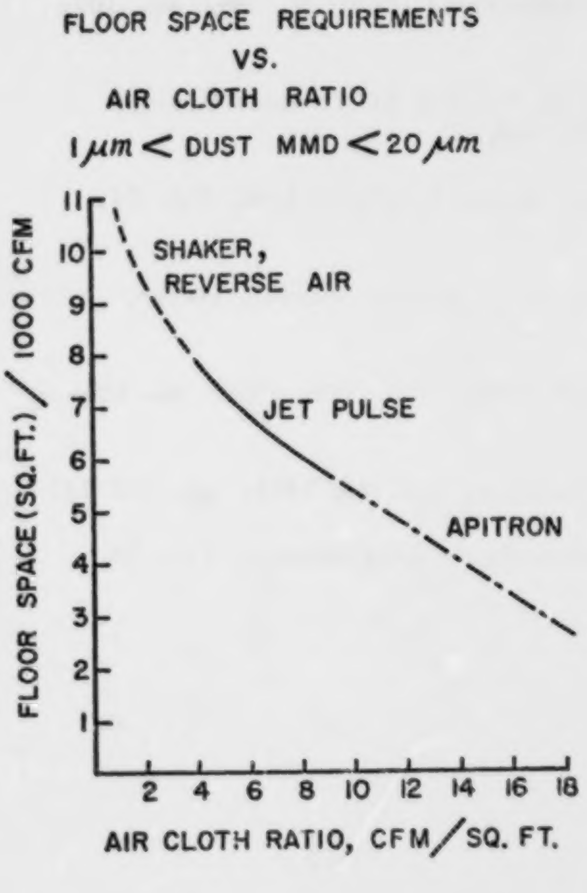

Figure (11) -

Typical range of air cloth ratios for three types of fabric filters, showing the resultant floor space requirements

The exact air cloth ratio required depends on the application, but, in general, when the ratio is in the high range for one type of filter, it will also be in the high range for the other two types.

It has been seen that the electrostatic augmentation of the fabric filtration process yields improved collection efficiency and reduced space requirements. The Apitron system can be regarded as a new category of particulate control device, offering performance unattainable with more conventional equipment.

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WRITTEN DISCUSSION

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The authors are to be congratulated for bringing to our attention a field tested filtration technique with much reduced pressure drop feature compared to a conventional bag-house design.

A central theme of the paper hinges on the gains achievable in lowering the overall pressure drop in the practical bag-house type filtration systems by the application of electric fields to such systems. This reduction is achieved with equal or better collection efficiencies when compared with a similar system of the non-charged case. Qualitatively it is asserted in the paper that the smaller pressure drop feature is due to dust cake being more porous in electrostatically charged systems and the increased efficiency is possible because collision cross section between the particle and the collector (fiber) is larger due to induced polarity.

Ramifications of the paper are enhanced further when the nation is turning to recognize the energy shortage and the conservation as an effective measure. The smaller pressure drop in a filtering system for the same flow rate and dust loading, but with equal or better collection efficiency, affords a direct reduction in the fan power necessary to drive the system. For the proposed system, however, economics should also be studied to correlate the cost aspect of the increased hardware and the maintenance of the fields. I presume the work being preliminary in nature will be expanded and encouraged to cover such matters in the future.

As noted earlier, data for the paper have been obtained essentially from a field tested pilot plant and to a limited extent from a bench scale model. Therefore, only qualitative discussions are possible. To increase accuracy and for comprehensive data, bench top models need be built. In this report, for example, a comparison of Figure 2 and 4 shows that at the same face velocity (15 FPM) and for the same cloth and dust variety (NOMEX/LIMESTONE) the data points shown in Figure 4 could be shifted to higher absolute pressure drop values resulting in less favorable pressure drop ratio between the non-charged and the charged filtration cases.

Examination of Figures 2,3 and 4 shows also that the pressure drop in the charged filtration system levels off after approximately 10 kV/in applied voltage gradient for all fabric and dust varities presently tested. Is the mechanism for such an applied potential gradient threshold well understood? Existence of such a "saturation" point is important in the system optimization for power reduction. Furthermore, if such a threshold, indeed, exists, then this value might be considered for normalizing the data.

The data presented in Figure 9 raises questions in two respects and the author's comments are welcome. First, is the given dust loading figure at outlet the same for both systems? Second, is indeed the single data point shown for the bag-house a valid data point? Or, were there some extraneous effects such as humidity difference between the two gas streams as the paper mentioned as a general comment.

Figure 10 is perhaps a more direct comparison of the improvements possible with a charged system over a similar non-charged system. It appears that even the non-charged APITRON has been a favorable design over an existing conventional bag-house system in terms of its pressure drop characteristics. Consequently, Figure 9 should be interpreted with this difference in mind.

In conclusion, while I find the authors' presentation most interesting and far reaching in its energy saving potential in engineering applications, I would like to emphasize the necessity for more comprehensive and well-controlled research in fundamental assessment and understanding of the system presented.

OPEN DISCUSSION

Helfritch: I would like to reply to Dr. Kuzay's comments.

First of all, I think we are referring to Fig. 2 and 4 in the manuscript. These figures show that at the same face velocity with the same flow and dust, the data points shown in Fig. 4 should be shifted to a higher absolute pressure drop. Those two sets of data were not taken under identical conditions. The one parameter that was different between the two was the relative humidity. This is also a factor that I have not talked about, but is being looked at. I think that accounts for the difference you see in pressure drop at given filtration rates.

Examination of Figures 2, 3 and 4 show also the pressure drop levels off at approximately 10 k volts/inch. I do not know if there is a threshold but it is possible. We could not usually go much beyond the voltages I have shown because that is the spark level point. If we were able to carry the voltage gradient out farther, I do not know whether we would reach some sort of a threshold or not. At the inlet the grains per cubic foot would be the same for both systems. What is shown in Fig. 10 is the outlet concentration.

Liu: Could you clarify the voltage gradient? Is this the voltage of the wire divided by the distance of the wire from the cylinder? Or is this actually the voltage of the wire?

Helfritch: It is not a good representation. The actual voltage gradient is not linear at all. It is just the average voltage.

Nichols: You showed a variation in pressure drop with applied voltage. Have you conducted any tests where you operated at a lower voltage than the corona inception?

Helfritch: No.

Nichols: There is a potential for the formation of dipoles that may lead to the same type of mechanism. If that is the case, you may want to put a pipe down the middle of your bag rather than having an auxiliary charger if only the dipoles are necessary. Secondly, have you made any attempt to determine the relative collection in the precipitator versus what it is in the bag filter?

Helfritch: On occasion we have done that and it varies quite a bit with the dust. The best I can say it is between 20 and 80 percent, depending on conditions and the type of dust we are using. It is not a very efficient precipitator. We run very high velocities through those pipes.

Nichols: I wonder if it is the charging of particles or whether it is the formation of dipoles that is important for the reduction in the pressure drop.

Helfritch: No, it is the charging of the particles.

Nichols: But it still could be forming a dipole that leads to the low pressure drop. They would not, necessarily, have to be charged, they could form dipoles.

Helfritch: There is no field to do that.

Nichols: How about from the previously deposited material?

Helfritch: But that requires a charged material.

Nichols: The charge is on there when you charge the particle. If all you need is to form a dipole, all you would need to do is form a field within the filter by having metal fabric in the bag with a pipe in the middle. If you form the dipoles at the slow face velocities, you would still have time for these to align and give the low pressure drop. I do not know whether it is the fact the particles are charged or the fact that they have formed the dipoles that leads to the low pressure drop.

Lamb: I have seen similar effects but we do not control the charge on the particles. We do not charge the particles, but apply a field in the plane of the fabric by just putting electrodes in the fabric itself.

Penny: Do you have figures on the resistivity of the dust at which this comparison was made? Precipitators are so dependent on this resistivity.

Helfritch: We have never measured resistivity. We have no equipment for that. We have noticed a strong effect that we interpret as high resistivity problems. We have also noticed strong corona quenching effects when we go to very small particles. In these cases, we generally have to run at lower voltages and currents than otherwise. However, even in these cases, particularly in Table 1 for the case of the cement kiln and the electric furnace fume, we were way down in voltage and current but we still got improvements in performance. I attribute low back voltage and current to resistivity of the dust.

Cooper: It may happen that changing the resistivity of the dust cake would change the particle size distribution. This might be expected if you have 20 to 80 percent collection in the electrical case beforehand. If you had a coarser dust that would be consistent with a lower pressure drop per weight per unit area on the filter. It also might suggest to some degree the different structure in the cake.

Helfritch: Again I allude to the bench scale-testing where none of the particulate was collected in the precipitator section. Dust particles were deposited in the charged state. Basically in our APITRON we are getting sort of the same thing.

Reif: With the configuration you are using, I do not think there would be any evidence to suspect that there are polarization effects present. Any particle polarized upstream probably would "unpolarize" by the time it got into the filter.

Nichols: The space charge would still have an electric field in the bag.

Helfritch: You are talking about a non-ionizing field; you are saying you produce a dipole before you produce corona. You do not have a charged particle so you do not have a space charge field.

Nichols: But you can have a field set up where you can induce dipoles to determine whether it is a charge or dipole that cause this effect. It is possible, but you did not do it.

In Figure 6, the dotted portion of the curve should look steeper. I would expect that there would almost be a vertical drop in pressure before the corona, unless there is something else going on.

Looking at Figures 2, 3 and 4, there is an area where there is no corona generated.

Grassel: Does this affect the cake structure? Does it affect the depth of deposit on the fabric filter and does it affect the strength of the agglomerate that comes off?

Helfritch: It does affect the cake structure. The depth of particulates penetrate into the filter would probably have to be reduced because most of the pressure drop you see across a felted filter is due to particles that have penetrated the filter.

ELECTROSTATIC CAPTURE OF FINE PARTICLES IN FIBER BEDS

D. L. Reid

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ABSTRACT

Experiments at Battelle-Northwest led to the discovery that charged particles could be efficiently collected by a bed of knitted fibers having a very high bed volume -- fiber volume ratio and very low pressure drop. The described concept depends on unipolarly charged particles being deposited on fibers of like charge. Results are presented for the system efficiency for submicron particles as a function of velocity, particle resistivity and bed depth and porosity.

INTRODUCTION

The capture of charged fine particles in highly porous fiber beds was discovered by A. K. Postma and W. K. Winegardner near the end of a study on the collection of charged submicron particles by oppositely charged spray drops. To demonstrate the enhancement of collection by the charged spray drops, the negatively charged particles were passed through a dry system which consisted of a spray chamber and, fortuitously, a polypropylene fiber demister. Surprisingly, a rather dramatic increase in the collection efficiency was observed rather than an expected lower efficiency. This observation inaugurated an investigation which eventually reduced the concept to practice and ultimately a patent.

Since the Demister® pad produced little or no attenuation of submicron particles without field charging, some phenomenon other than natural deposition processes was involved. Subsequent tests revealed that the collection mechanism depended on the development of an induced space charge within a bed of electrically non-conducting fibers. This is simply accomplished by passing the aerosol through a negative corona field and then through the fiber bed. The deposition of charge on the fibers instantaneously creates a negative space charge field within the bed. Then if a highly electrically resistive fiber is used and the particle is electrically non-conducting, the bed field charge is maintained at a magnitude sufficient to produce a high removal efficiency.

THEORY OF CONCEPT

The mathematical modeling of the concept, by D. L. Lessor, is too extensive for this presentation so only a brief explanation is presented. Since the electrical field in the bed is induced by the deposition of charge, efficient operation of the system depends on the development of electrical fields of like polarity sufficient to overcome the coulombic repulsion force of a single fiber within times shorter than the residence time of the particle within the bed.

As the charged aerosol enters the bed of like charge, the interaction between one particle and the fiber nearest to it is a repulsion force. However, the charge distribution on the whole bed produced an electric field force that drives the charge from the bed. This bed scale field apparently has a force sufficient to divert the charged particle from the normal flow lines and onto the fibers. Thus, the fiber bed concept is effective when the bed scale field force is significantly greater in magnitude than the repulsion force of a single fiber. The following calculation compares the magnitude of the bed scale field in the gas flow direction (z), E_z, with the radial electric field, E_r. caused by the nearest fiber.

Consider a fiber filter of thickness 2h in the direction of the gas flow and infinite in the transverse direction. If the fiber length per unit of pad volume, ℓ_v, is uniform and the fibers carry uniform charge per unit length, q_ℓ, the macroscopic charge density, ζ, is

$$\zeta = \ell_v q_\ell$$

This charge density serves as a source for the bed scale electric field whose local average over several fiber spacings has a component E_z given by

$$E_z = \frac{q_\ell \ell_v z}{\varepsilon} \quad (\text{for } -h \leq z \leq h)$$

where ε is the permitivity of the bed, and z is measured from the midplane of the bed. Positive values are in the direction of the gas flow.

The radial electric field from the nearest charged fiber is

$$E_r = \frac{q_\ell}{2\Pi r \varepsilon}$$

where r is the distance from the center of the fiber.

Then the ratio of these forces at the fiber surface is

$$\frac{E_z}{E_r} = 2\pi \ell_v r_f z = 2\beta \frac{z}{r_f}$$

where r_f is the fiber radius and β is $\pi r_f^2 \ell_v$, the fraction of the bed occupied by fibers.

This ratio is much larger than unity for the conditions used in the fiber bed concept. For example, with an occupied bed fraction of 0.05 and a ratio of bed thickness to fiber radius of 500, then E_z/E_r is 50.

A computer program has been developed which calculates steady-state conditions for simultaneous charged aerosol deposition, charge conduction on the bed, charge leakage off the bed edges, and effect on deposition by the field from the retained charge. This modeling work indicates that

- electrically resistive beds retaining substantial charge provide enhanced filtration efficiency and
- the bed scale electric field effect is a dominant collection mechanism for submicron aerosols.

EXPERIMENTAL PROGRAM

Based on initial observations variables thought to influence system efficiency were (1) particle-size, concentration, resistivity and charge level, (2) gas-velocity, temperature and composition, and (3) fiber bed-resistivity, thickness, porosity and dust surface coverage. A complete block experimental program involving all variables would require more than a practical number of tests for proof of the concept. Therefore, the initial program was designed to look principally at the bounding conditions over some reasonable operating range. Table I lists the test conditions for the parameters of interest.

TABLE I. System Test Conditions

Parameter	Range
Particle Size	0.06 - 3 μm
Aerosol Concentration	10 - 250 mg/M^3
Particle Resistivity	10^8 - 10^{13} ohm-cm
Particle Charge Level	Near Saturation
Gas Velocity	0.25 - 2 m/sec
Gas Temperature	20 - 130 C
Gas Composition	Air and Combustion Products
Fiber Resistivity	Conducting and Nonconducting
Bed Thickness	7.5 - 30 centimeters
Bed Porosity	Void Fraction - 0.94 - 0.98

TEST SYSTEM

Figure 1 is a schematic drawing of the system which consists of a 7 foot long 24 inch diameter pipe, a short corona charging section with an effective cross-sectional charging area equal to that of the duct, a 3 foot long section between the charger and fiber bed to preclude any direct field charge effects from the charge section, and a bed frame for the knitted fiber mats. The inlets and outlets to the bed were 24 inches diameter. Since the bed was hand packed, the bed face dimension was made 28 inches by 28 inches to extend the experimental run time before edge leakage became the predominant source of downstream particles. The frame was made with lucite to accommodate up to a 12 inch thick bed and was electrically isolated from the rest of the system as was the charging section. The system was built in sections for future alteration of configurations and distances between principal components and was held together with electrically insulated external straps. A dispersion plate was placed in the 12 inch diameter inlet at the transition point to produce the normal velocity profile at the three principal sampling points at positions 1, 2 and 3.

FIBER BED VOID FRACTION

The void fraction of the fiber bed was defined as that volume of the bed not occupied by the fiber volume and was determined from the bulk weight of the fiber mats, density of the fiber per unit length and the fiber diameter. The knitted mats were supported within the frame by a grid of teflon coated wires having about a 4 inch wire spacing. The 15 cm thick bed was packed to ~0.96 void fraction (VF). The 7.5 cm bed had an ~0.94 VF. For the 30 cm thick bed, the first 15 cm had a VF of ~0.975 and the last 15 cm ~0.96. The lower fiber density in the first half was used to increase depth of penetration and thus increase the load to bed pressure drop ratio. This accomplished both goals, but appeared to be detrimental with respect to efficiencies in the later stages of a run series as well as length of operating time before a decrease in efficiency was observed. Visually, the depth of penetration appeared to be directly related to particle resistivity for the submicron aerosols -- increasing with increasing resistivity. Particle size was not a primary variable. Interest was in the 0.1 - 0.3 μm size range except for fly ash which was collected downstream of a two-stage ESP.

PARTICLE GENERATION AND CHARACTERISTICS

Four particles were used in this series of experiments to highlight any change in efficiency with particle resistivity -- namely, amonium chloride, magnesium oxide, sodium oxide, and fly ash. The fly ash, obtained from a plant burning low-sulfur Western coal, was resuspended by aspiration. The NH_4Cl particle was generated by air sparging of HCl and NH_4OH solutions with interception of the gas streams at the inlet to the system. The particle generation rate was controlled by varying both the concentration of the solutions and the sparge air flow rate. The MgO and Na_2O particles were generated by metal vaporization and oxidation. The metal was heated inductively in a boat enclosed in a tube. An inert gas forced the metal vapors

WIND TUNNEL SCHEMATIC

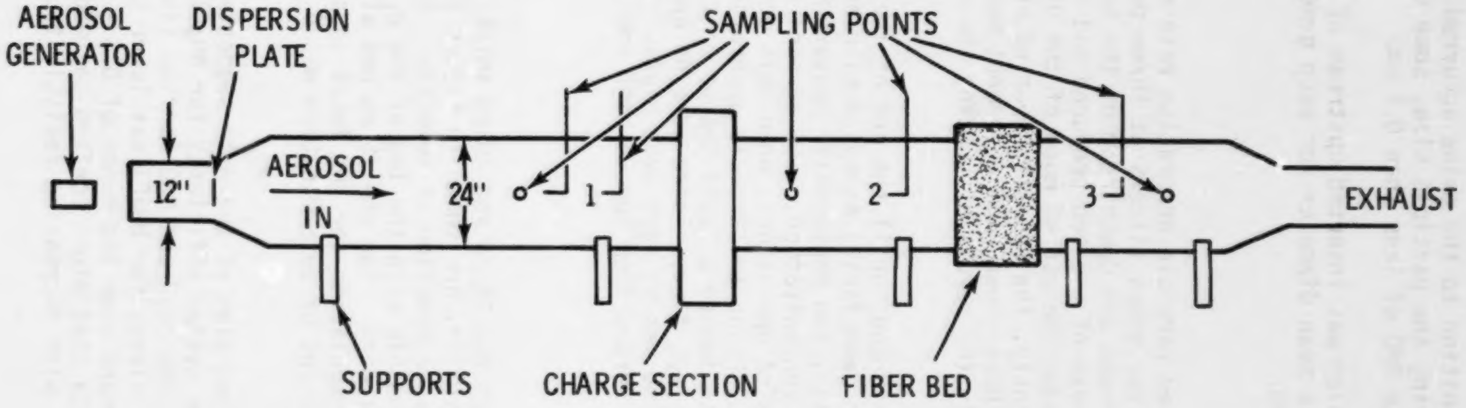

AEROSOL GENERATOR
DISPERSION PLATE
12"
AEROSOL IN
24"
SAMPLING POINTS
1
2
3
EXHAUST
SUPPORTS
CHARGE SECTION
FIBER BED

FIGURE 1

out the end of the tube where ignition to the oxide occurred. Since a technique was available for varying the particle size, some tests were made with MgO particles having an AMD of less than 0.1 μm.

An eight-stage Anderson sampler was inserted upstream of the charge section to obtain the aerodynamic mean diameter for each generation in which the conditions were altered.

TEST PROCEDURE

After establishing the desired particle generation rate mass samples were collected on membrane or fiber glass filters at three positions: upstream of the charger and upstream and downstream of the bed. The high particle and field charges upstream of the bed produced odd results that defied interpretation and precluded the use of much of the data for determining bed efficiency. Consequently, the results reported are for the system efficiency. Later tests have shown good agreement between bed and system efficiencies -- apparently the result of cutting the current flow in the charge section by one-half.

Generally, operation was continuous until the end of the day or until apparent breakthrough (whichever came first) except during adjustment of generation rate and flow. Usually, two consecutive measurements were made before changing the flow and/or concentration. Low aerosol concentration runs generally required two days of operation. When this occurred, the system was shut down over night. Prior to particle generation the second day, the charger was turned on to impose a field charge on the bed and a reentrainment test made at 1.5 m/sec flow. There were no apparent degrading effects even though the bed charge was at zero over night. For a series of measurements, the flow rate was varied from low to high and back to low or in the reverse order.

RESULTS AND DISCUSSION

Typical results obtained with the 15 cm and 30 cm thick beds whose void fractions were described earlier, are shown in Figure 2. The decontamination factors are plotted as a function of velocity. The data approximate the expected relationship with the log of the D_f varying linearly with a power of the velocity. For the 15 cm bed all data were considered to be from the same population for the least squares fit represented by the line. The coefficient of determination was 0.9 and all data fell within the 2σ error limits.

For the 30 cm bed, the log-log plot of the data suggested a real although small difference in the system efficiency for NH_4Cl particles. The NH_4Cl data were segregated from Na_2O and MgO for the linear regression analysis. The average efficiency for NH_4Cl was lower than that of MgO or Na_2O by about 2 to 5 percent over the range of 0.5 - 1.5 m/sec face velocity. This is the first data that might confirm the model's prediction that efficiency should decrease with decreasing resistivity. Since the

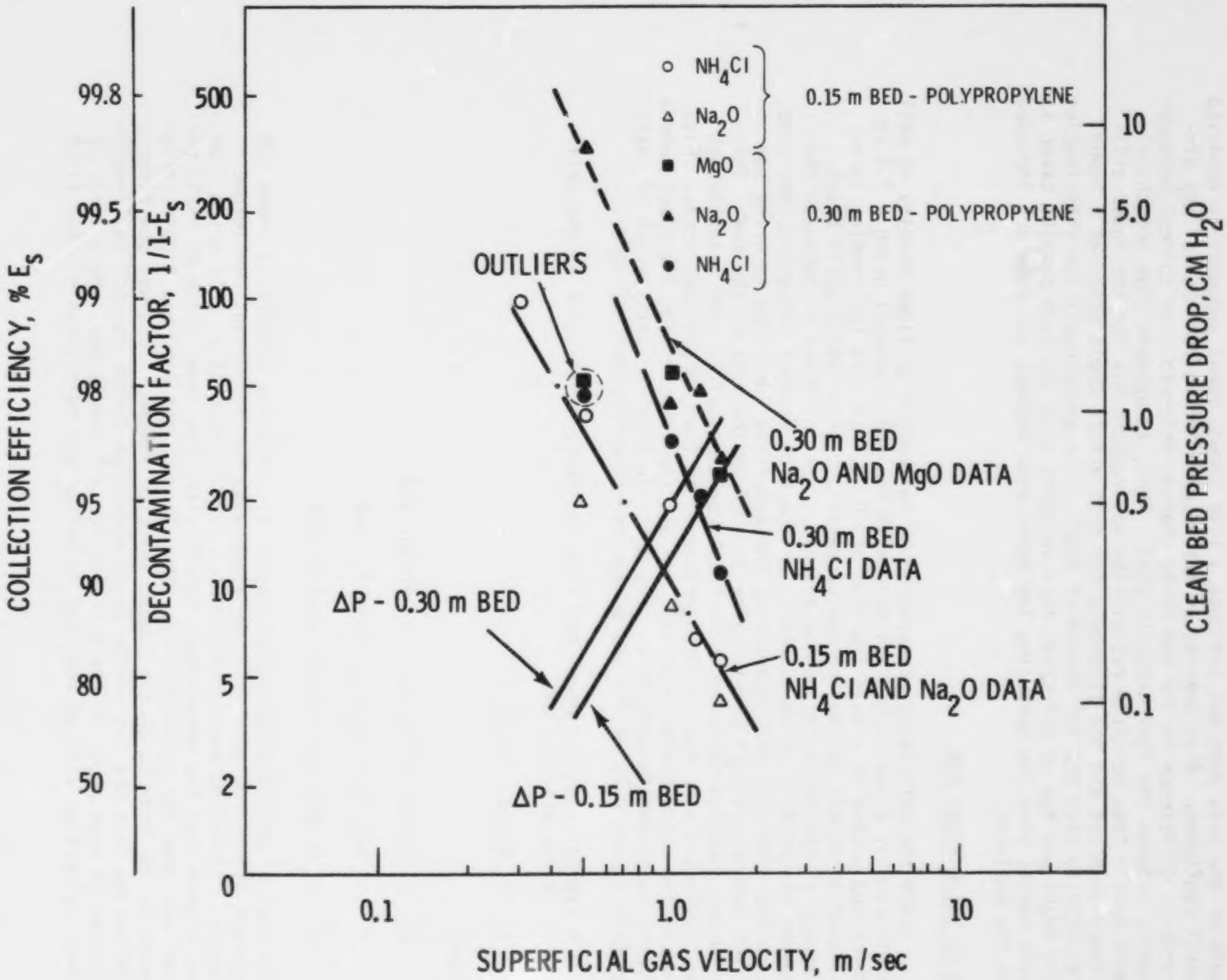

COLLECTION EFFICIENCY, % E_s
99.8
99.5
99
98
95
90
80
50
DECONTAMINATION FACTOR, $1/1-E_s$
500
200
100
50
20
10
5
2
0
CLEAN BED PRESSURE DROP, CM H_2O
10
5.0
1.0
0.5
0.1
SUPERFICIAL GAS VELOCITY, m/sec
0.1
1.0
10
○ NH_4Cl
△ Na_2O
0.15 m BED - POLYPROPYLENE
■ MgO
▲ Na_2O
● NH_4Cl
0.30 m BED - POLYPROPYLENE
OUTLIERS
0.30 m BED Na_2O AND MgO DATA
0.30 m BED NH_4Cl DATA
0.15 m BED NH_4Cl AND Na_2O DATA
ΔP - 0.30 m BED
ΔP - 0.15 m BED

FIGURE 2

bulk of the data does not indicate a like trend, final judgment is reserved until confirmed. Also bearing on the reservation is the lack of any discernable difference in the bed field charge, inferred from current measurements, between the four aerosols used. Thus, it appears that something more subtle than particle resistivity was responsible for the lower efficiency observed and the threshold for this effect would occur at a lower resistivity than the 10^8 ohm-cm of NH_4Cl. The anomalously low efficiencies for NH_4Cl and MgO at 0.5 m/sec face velocity were the last samples taken in each series when the operating log noted edge leakage and were not included in the analysis.

8 cm Thick Fiber Bed

Since the efficiency appeared to be influenced by fiber density as well as thickness, a run series was made with the NH_4Cl aerosol using a 7.5 cm thick bed packed to a void fraction of ∿0.94. In this test series it was planned to obtain successive measurements at one velocity until breakthrough at the bed edges became apparent. In an effort to increase the number of consecutive measurements at the high aerosol concentrations used, a folded fiber mat was inserted at the bed-frame interface to bring the fiber density at that point nearer to that in the rest of the bed. This produced a marked improvement in total run time before edge breakthrough was indicated at a much higher pressure drop than in previous tests. Five successive measurements were made for all flowrates except 0.5 m/sec where twenty measurements were made. The particle size for this series of tests hovered around 0.25 μm AMD and a mass concentration near 75 mg/M^3.

A typical equation developed from the least squares fit of the efficiency versus velocity was

$$E_S = 102.8 - 10.28\ V$$

where

E_S = system efficiency in percent and,

V = bed face velocity in m/sec and

r^2 = 0.986, correlation coefficient.

The plot of the decontamination factor vs. velocity (Figure 3) shows the average values for each velocity along with the 95 percent confidence levels. Since the average efficiency (99.95 percent) at 0.25 m/sec flow was higher than the measurement sensitivity, the lower value of 99.93 percent was used for curve fitting. The slope of the line was highly influenced by the high D_f at the lowest velocity tested. This was the first indication that the log of the decontamination factor may not be linearly related to some power of the velocity for all conditions. More tests at lower velocities will be required to define the relationship. But first

DECONTAMINATION FACTOR
7.6 cm POLYPROPYLENE BED
0.946 VOID FRACTION

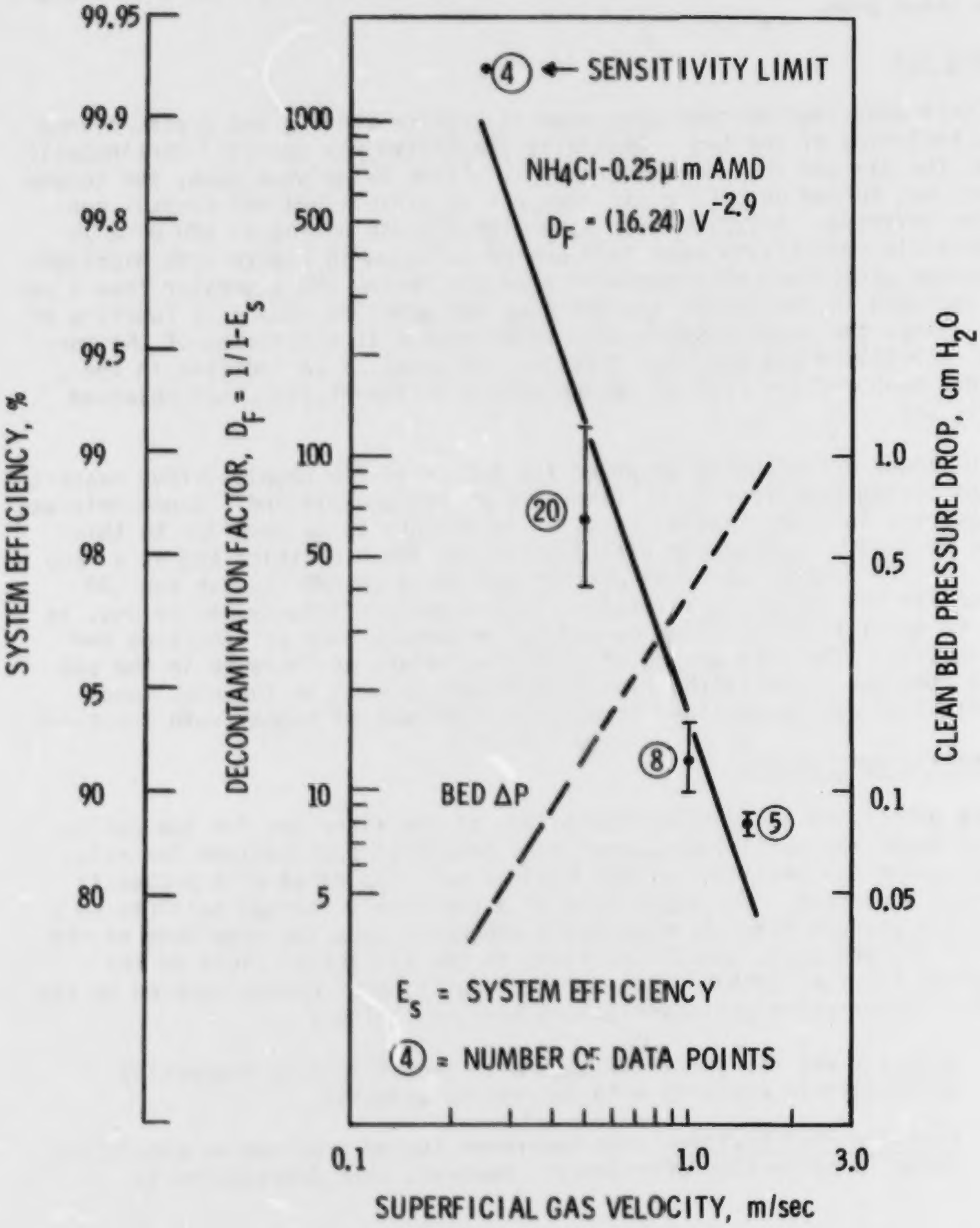


FIGURE 3

some measurement method other than mass, if available, will be required to provide adequate sensitivity at the lower velocities.

Comparatively, this thinner ~8 cm bed with a lower void fraction exhibited an efficiency about equal to the thicker beds at 0.5 m/sec flow and at 1.5 m/sec flow was about 5 percent higher and lower than the 15 and 30 cm thick beds.

Loading Run

Continuous loading runs were made to examine loading and pressure drop characteristics of the bed. Generally the system was operated continuously during the day and shut down over night. After being shut down, the corona charger was turned on before air flow was re-established and aerosol generation effected. A typical run made with fly ash having an AMD of 2 μm and particle resistivity near 10^{13} ohm-cm is shown in Figure 4 to highlight the higher efficiency obtained with aerosols having AMD's greater than 1 μm. Also included in the figure are the load and pressure drop as a function of time. Since the load-pressure drop relationship is a function of the particle resistivity and bed void fraction the onset of an increase in the pressure drop and the rate of change may be different than that observed for fly ash.

The lower efficiencies observed for the first few samples after restarting the system the first five times are as yet unexplained. Since this was not observed in other loading runs, it is thought to be peculiar to this series of tests. Comparable efficiencies for the 8 cm thick bed at a face velocity of 1.5 m/sec were >99 percent for the 2 μm AMD fly ash and ~90 percent for the submicron aerosols. This higher efficiency may be due, in part, to particle attenuation by natural processes such as impaction and interception. The long period of operation before an increase in the bed ΔP was observed reflects the high void volume as well as the high load capacity that can be obtained especially with beds of higher void fractions.

SUMMARY AND CONCLUSIONS

The efficiency and design flexibility of the fiber bed for the collection of submicron particles suggest many practical applications for relatively economical particle control that is not attainable with presently available equipment. The deposition of a negatively charged particle on a negatively charged fiber is apparently dependent upon the magnitude of the ratio of the bed scale electrical field to the electrical field of the individual fiber -- both created by the deposition of charge carried by the aerosol. Other principal observations were as follows:

- Velocity was the prime variable with the efficiency decreasing approximately linearly with increasing velocity.
- Particle resistivities over the range tested produced no consistent trend in collection efficiency. However, some degradation in

POLYPROPYLENE FIBER BED
7.5 cm THICK - 0.94 VOID FRACTION
FLYASH - 2μm AMD - 10^{13} OHM - CM RESISTIVITY
1.5 m/sec FACE VELOCITY

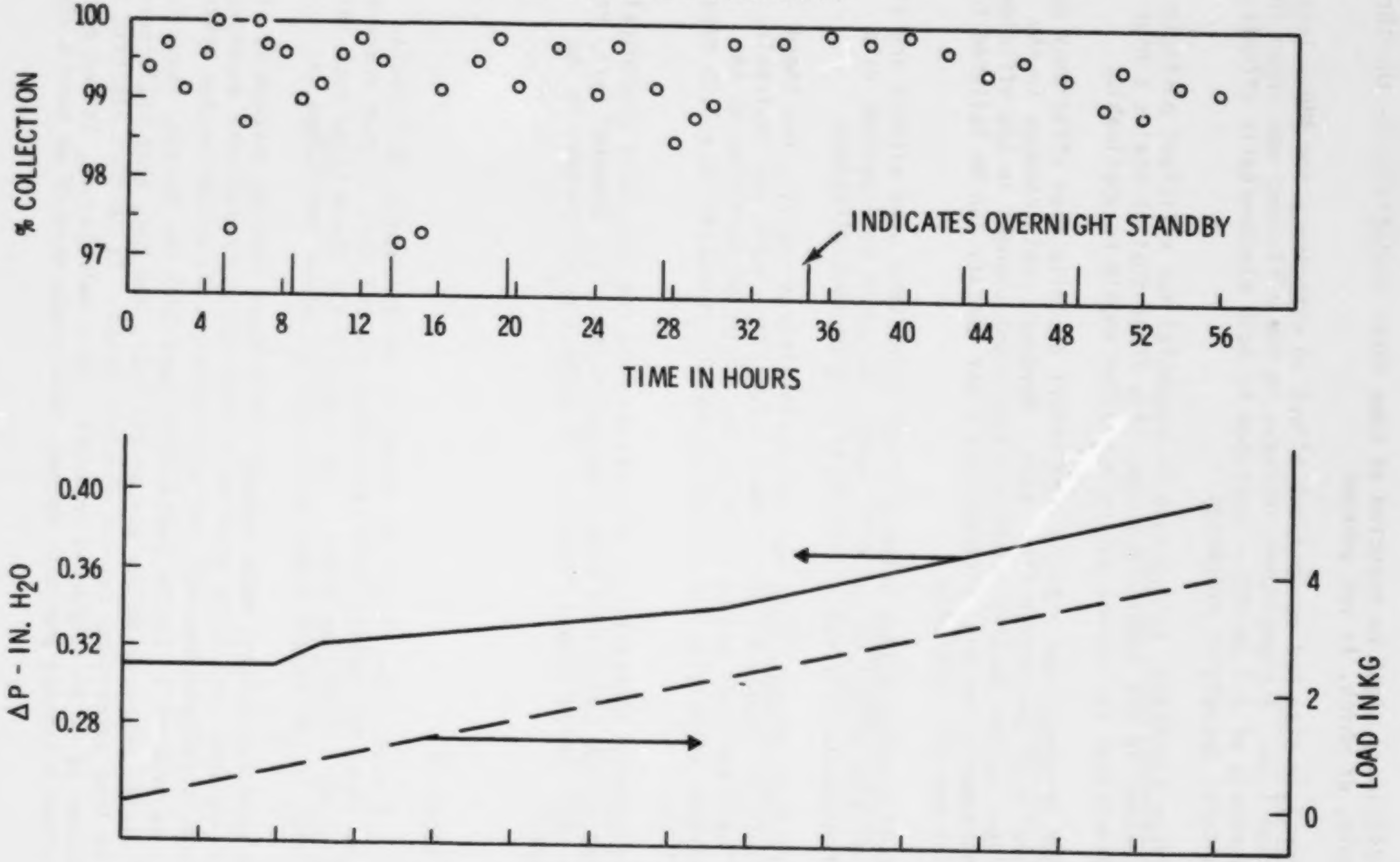


FIGURE 4

efficiency might be expected at some lower resistivity -- the threshold, of which, is yet unknown.

- Particle size had no apparent effect on efficiency for AMD's less than 1 μm. A significant increase in the efficiency was shown for aerosols of 2-3 μm AMD -- ascribed to both electrostatic effects and natural deposition processes.

- Fiber electrical resistivity is essential for efficient particle control in the submicron range. Any fiber that maintains a high electrical resistance during operation should be applicable.

- Bed thickness had a pronounced effect on collection efficiency especially at the higher flow rates. However, small changes in the fiber volume density produced significant changes in the efficiency. Apparently the bed thickness and fiber density can be tailored for the aerosol condition.

- Dust concentrations between 10-250 mg/M^3 were used without any significant effect on bed efficiency. As might be expected, data reproducibility was better at the higher concentrations.

- Particle charge level was not varied intentionally. For these initial investigations, a near saturation charge was desirable. The fiber beds had no affinity for uncharged particles at the higher velocities and showed ∿8 percent retention at a 0.25 m/sec flow.

The test results substantiate the influence of the bed scale electrical field forces for collecting submicron particles. The computed E_z/E_r ratio for the beds tested ranged from 20 - 70 without major changes in the efficiency.

OPEN DISCUSSION

Melcher: I would like to set the trend of the discussion if I could. I use a rule of thumb in dealing with electrostatics and that is that there is no such thing as a repulsion force. The reason I say there is no such thing as a repulsion is you really ought to account for where the charge is.

You have the filter, with charged particulate passing through it, the question arises: how is it that the collection sight collects anything? Let us look at a calculation, call the half spacings, S, my contention would be that one way to look at it is the particulate gets into the matrix, then, in a van de Graffe sense, with the gas as the belt, and this (the collectors) as the resistic load generating its own electric field. It is that electric field that then allows it to be collected on sight. As a calculation, let us say we have μm as a density per unit volume, that would mean if we have a charge on each of the particles in there, we would have a charge per unit volume which would be q.

Gauss' law says that coming out of this region is an electric field which is equal to S times the charge density, or $\varepsilon_o E$ actually. So if you look at a particle out in front we find that it is experiencing this sort of polarizing field. Not only does it have a charge, q, but it is also in an ambient field. That means the particle with charge q is generating a field that would equal $R \frac{q}{4\pi\varepsilon_o}$, where R is the radius of this spherical particle. So the situation is such that the particle experiences the ambient field and out of the particle comes its own field. Certainly if plus particles are going to get collected on that surface, we have to have E lines coming in. So you see how you might solve the problem. Just look at it as a conducting sphere in a uniform field carrying that charge. You would find that it is a matter of whether there are any E lines coming in. It depends on if $(\frac{q}{4\pi\varepsilon_o} R)$ is less than the E field coming in, which is $\frac{3\mu qS}{\varepsilon_o}$ where 3 is the concentration factor. You can then cancel out the charge and ε_o. Let us also view the packing as being determined so we get the particle size as much as possible by the spacing between particles, αR. In other words I am normalizing the spacing to the size of the particle. In that case the number density is $\mu = \frac{1 \text{particle}}{\alpha^3 r^3}$. We make the condition that the particles will collect if

$$1 < \frac{(3S)(4\pi)}{R\alpha^3} .$$

What this tells you is that the thickness of the filter must be bigger than the size of the fiber particles, which is a very good approximation. In order for it to compete with an electrostatic precipitator, you want this self-generated electric field to be larger than what you have in the ESP. It will do much better in resonance time than ESP. To me it is an electro-packed bed with a self-generating field, where this is a big resistor that drains off current. One thing that bothers me about the experimental work is how much it can be related to something else when in fact you are not held accountable for where the images are. If the load is on the edges then the field is not actually one dimensional, it is actually quite complicated and will depend on the size. If you had screens clearly the field would be longitudinal and you would have the thing well under control. I would worry how relevant it is to scale-up.

There is a patent by Cole in which he packs between screens particles which are semi-insulating. He sends the current through and collects in this mode. What he has done is to apply the field rather than to self-generate it. That would be my view of it.

Reid: I did not do the mathematical modeling. This was done by D. L. Lessor. This is true if you have a uniform charge throughout the bed and a uniform fiber length per unit volume. At this position in the bed plane, essentially your charge effects are zero. However, this is not the condition encountered in the fiber bed. If you look at it with respect to this and you look at the ratio of electrical force in the bed versus the electrical force at the surface of the fiber, you end up with a relation of $\frac{2\beta Z}{RF}$. These are just physical dimension of the system but it does give an approximate relationship of the bed scale field forces with respect to the force at the fiber surface itself. The system, as we have run it, varied from ratios of 3 to 10. We interpret the repulsive forces as acting on the particle itself and diverting them from the normal flow lines that you would expect in the system. That would mean a diversion of submicron particles from their normal flow lines to allow them to deposit on the fiber itself. The magnitude of the bed scale field effect overcame the repulsion forces of the fiber and allowed it to deposit.

Melcher: This, by the way, is the configuration in which Whipple & Chalmers say an isolated particle will not collect. This is that "backward configuration." One that we say collects just as well. But one had better put something into his model that adds turbulence, or has something about the wake in it.

Reid: He has a very extensive mathematical model development of this system. It is worthy of a paper in itself and I hope it will be published in the not too distant future.

On the upstream face of the 12 inch bed, it appears that it is a completely closed structure. It should have an extremely high pressure drop in this system. Under these conditions the pressure drop at the time we terminated this test is about 1.2 inches of water. The particles deposited completely around the surface area of the fiber not in any one particular orientation, front or rear. The initial modeling work predicted the particles would all deposit on the rear face of the fibers. He had to revise his mathematical modeling on the basis of his observations. The particles deposited around the fibers.

Cooper: The thing that intrigued me most was that if you shut it down, waited overnight, turned the gas on the next day with a corona charger, you could get the thing back up to the condition where it was catching charged particles. That suggests to me that something other than the particles themselves was bringing the charge to the fibers.

There are two possible solutions: 1) You are some how getting it from the ions and 2) Maybe we are seeing some sort of tribo effect in the plastic. That fact really puzzled me.

Reid: If we run on uncharged particles for the bed we see a positive leakage current, not a negative one. That suggests that the bed is producing a positive charge. It is very weak charge we are producing under the hydro-electric effect, principally since we have such an open structure.

It is true that all we had to do is turn on the charged section, although we could not use filtered air, and all air has particles in it. So whether we are looking at ion deposition or charged condensation nuclei count particles that are deposited on the fiber, we do not know. We produced approximately an order of magnitude less field force in the bed than when we were generating particles. We do not know whether it is ion deposition or deposition of condensation nuclei that have been charged. We know the system is about 95% efficient for condensation nuclei particles.

Cooper: I am still puzzled on how a negatively charged particle is deposited on a negatively charged bed. The strong coulombic repulsion, it would seem, would not allow a deposition.

Reid: The key is the ratio of the field forces.

Liu: I would like to suggest another point. Your device is very similar to a space charge precipitator,except you have localized, immobilized centers on which the space charge can be held stationary. So your figures depicting repulsion are still valid except that in half the electric field you have additional condensation centers which would act as impaction centers for the particle. For the case of the space charge precipitator, the particle would have to be driven all the way to the driving surface before it could be collected. It is a combination of Melcher's fluidized bed device plus the space charge precipitator.

Reid: That's pretty much right.

Friedlander: I am a little confused. Where is the energy coming from?

Melcher: It is a self-induced space charge. It is a gaseous van de Graffe.

Reid: The particles are the source of the energy. There is an imposed charge put on the particles upstream of the system.

FINE PARTICULATE AND GASEOUS EMISSIONS CONTROL EXPERIENCE WITH THE TRW CHARGED DROPLET SCRUBBER

R. R. Koppang

TRW Inc., Energy Systems Group

ABSTRACT

A unique, hybrid electrostatic precipitator for the collection of particulate and gaseous emissions has been developed and commercialized by TRW. The TRW Charged Droplet Scrubber (CDS) combines high energy wet scrubbing collection for micron size particulate with electrostatic precipitation for submicron particulate collection. The gas cleaning process steps of water droplet atomization, acceleration, particulate collection and demisting are performed through applied principles of electrostatics within a gas treatment equipment length of eight feet. Other CDS operating advantages include: no moving parts in the gas stream; low sensitivity to particulate physical properties; and low liquid-to-gas ratios. Particulate collection efficiencies exceeding 99% and 96% have been accomplished in the micron and submicron region, respectively. HCl gaseous emissions control from 90 to 99% have been obtained in pilot scale work with minor modification of the equipment internals. SO_2 flue gas desulfurization also appears promising.

INTRODUCTION

Recent, more stringent air pollution controls require high collecting efficiencies of submicron particulate to meet stack opacity regulations. The simultaneous control of gaseous and particulate emissions with one unit operation is highly desirable. Such collection must be accomplished with equipment which minimizes capital costs and significantly reduces energy consumption.

The CDS employs a unique engineering concept which provides for more diverse and economic applications of electrostatic precipitators in particulate emission control. Both large and small gas flow rates can be economically controlled with modular preassembled CDS equipment.

The CDS is a hybrid particulate collector using two parallel mechanisms; high energy water scrubbing and electrostatic mechanisms. These mechanisms are further augmented by the gas conditioning effects of droplet evaporation. Larger particulate is captured through high energy collisions with optimally sized water droplets. The benefits of the small equipment size associated with wet scrubbers and the low energy requirements of electrostatic atomization and precipitation are thus obtained.

The CDS is less sensitive than ESP's to particulate physical properties. The electrostatically sprayed water droplets create highly concentrated space charge fields and the evaporative losses are sufficient to improve the stability of high voltage breakdown. The walls of the plates used to collect charged particulate and demisted charged water droplets are evenly wetted and the material continuously flushed as a slurry. Its small size minimizes plant modifications, where space is usually limited, and permits factory assembled equipment to 100,000 cfm capacity. Low liquid to gas ratios, less than 1 gal/1000 cfm, minimize waste water treatment and the operating and capital costs associated therewith, and frequently allow waste water to be returned to other points in the factory production processes without secondary treatment. In many applications the CDS discharge gas is not saturated thus allowing less expensive fan and duct materials and minimizing the frequency of steam plumes.

These attributes of the CDS have found wide appeal among users in a number of basic industries, such as pulp and paper or iron and steel, as well as several specialty applications.

Described below are some of the more unusual operating principles. This is followed by a discussion of recent development and commercial experience with equipment configured for improved performance on submicron particulate and gaseous emissions. A previous paper, Reference 1, reported on earlier pilot and commercial experience on a first generation design.

OPERATING PRINCIPLES

The CDS operating principles are briefly summarized below. More detailed explanations may be found in References 1 and 2. The CDS accomplishes the major process steps for particle collection simultaneously within a small treatment zone using electrostatic principles for droplet atomization, charging, acceleration, particulate collection and demisting.

In the use of electrostatic atomization the CDS is unique to chemical process equipment design. Electrostatic spraying results in well dispersed water droplets which, when collected, evenly distribute water films on collector plates. Within the CDS water distribution and atomization are accomplished by an array of metering tubes arranged along a conducting, water carrying, high voltage energized electrode, Figure 1. Local, high voltage gradients are created at the tips of the tubes located equidistant between grounded parallel collecting electrodes (plates).

During the engineering development of the CDS various types of electrostatic atomizer designs, hollow tubes and rods, were considered. Currently a tube design has been selected which provides sufficient water flow rate for good droplet spatial distribution, optimal mean droplet size for high collecting efficiency, good plate water film coverage, and acceptable evaporation rates. High speed (1 second) photographs of the electrostatic spraying phenomenon for a voltage gradient of 5 kv/cm equivalent to corona breakdown in air are shown in Figures 2 and 3. Without high voltage, the water jets issuing from the tubes do not become atomized because of the low feed pressure (under 10" wc). As voltage is increased above 2-3 kv/cm, the water column becomes highly unstable bursting into a convoluted filament as it searches for a minimum gradient within the space charge field. A highly filtered DC (±5%) power supply is required to obtain stable electrostatic spraying at maximum operating voltage. Electrode operating rms voltages (35-55 kv) are typical of conventional dry electrostatic precipitators having similar electrode geometry.

Figure 2 is a time sequence photograph (15 exposure) which recorded the position of the water column as it adjusted to the droplet space charge; from this photograph the excellent space distribution of water droplets can be noted. The Figure 3 photograph recorded the early stages of disintegration of a portion of filament that separated from the main column; charged droplets are sprayed off the ends of the column and from kinks in the middle. Filament break-up processes, including local surface separation where surface charge forces exceed tension forces (Rayleigh instability), can also be observed in both pictures. Typical droplet sizes were in the range from 120 to 180μ m model (geometric) diameter, and 300 to 400μ m mass-mean diameter. Droplet size is influenced by nozzle tip geometry and water flow rate. The maximum velocity component seen for both 120μ droplets and 180μ droplets was about 30 m/sec.

Altering the physical properties of the spray liquid by adding a surfactant to reduce surface tension or soluble salt to increase conductivity have only secondary effects on spray patterns.

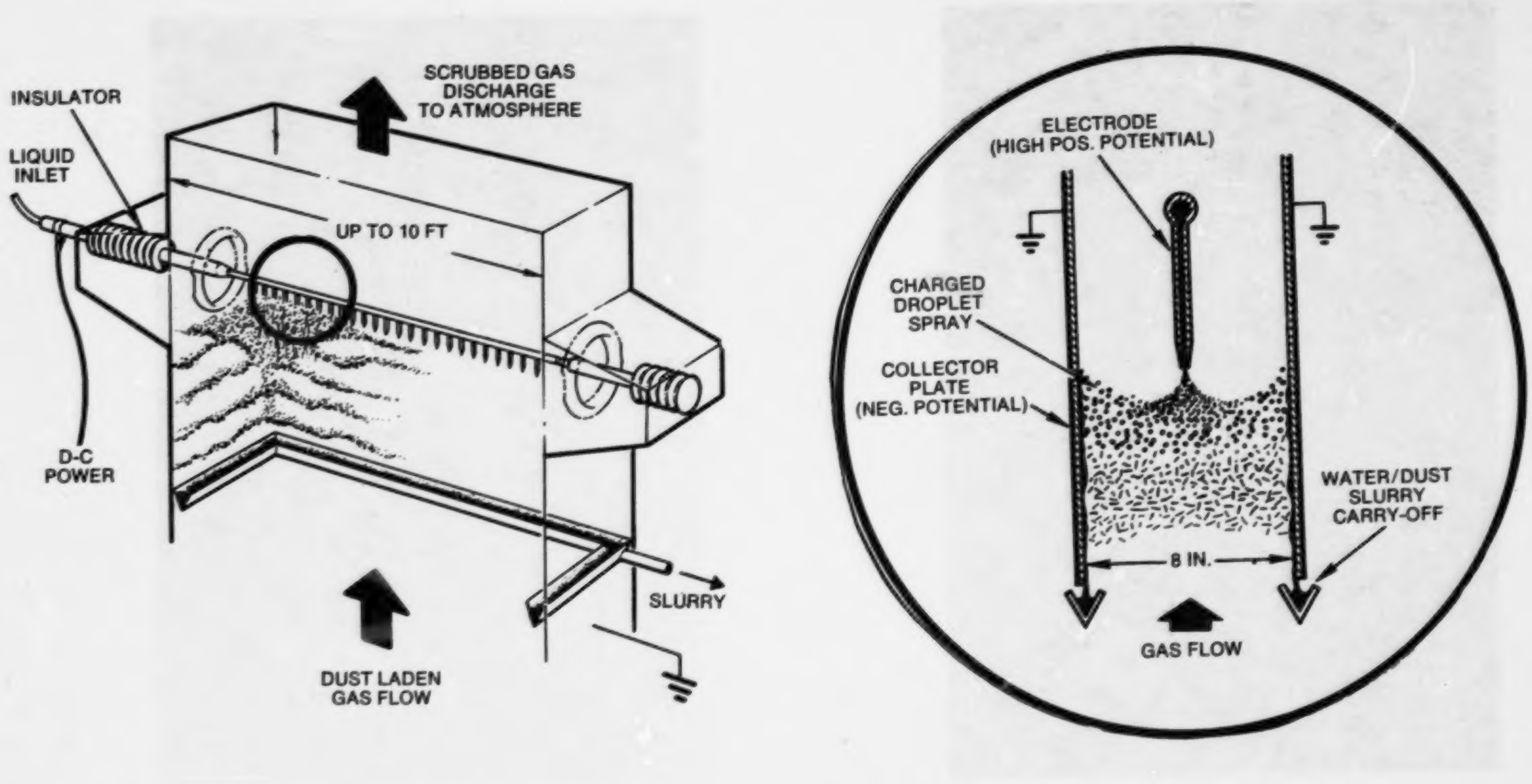

SCRUBBED GAS DISCHARGE TO ATMOSPHERE
INSULATOR
LIQUID INLET
UP TO 10 FT
D-C POWER
SLURRY
DUST LADEN GAS FLOW
ELECTRODE (HIGH POS. POTENTIAL)
CHARGED DROPLET SPRAY
COLLECTOR PLATE (NEG. POTENTIAL)
WATER/DUST SLURRY CARRY-OFF
8 IN.
GAS FLOW

Figure 1 Charged Droplet Scrubber Operating Principle and General Component Arrangement for a Single Electrostatic Spraying Stage

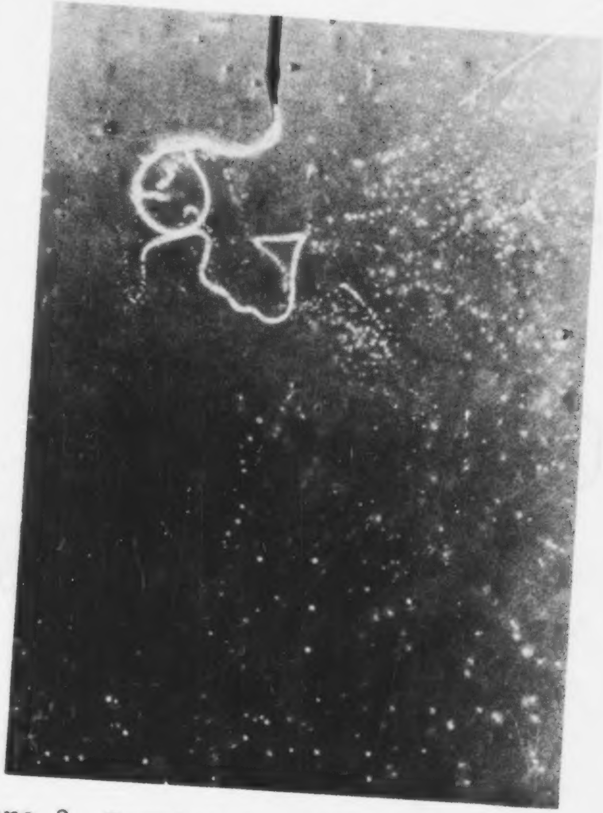


Figure 2 Droplet Formation by electrohydrodynamics spraying, expsoures at 1/15 second intervals

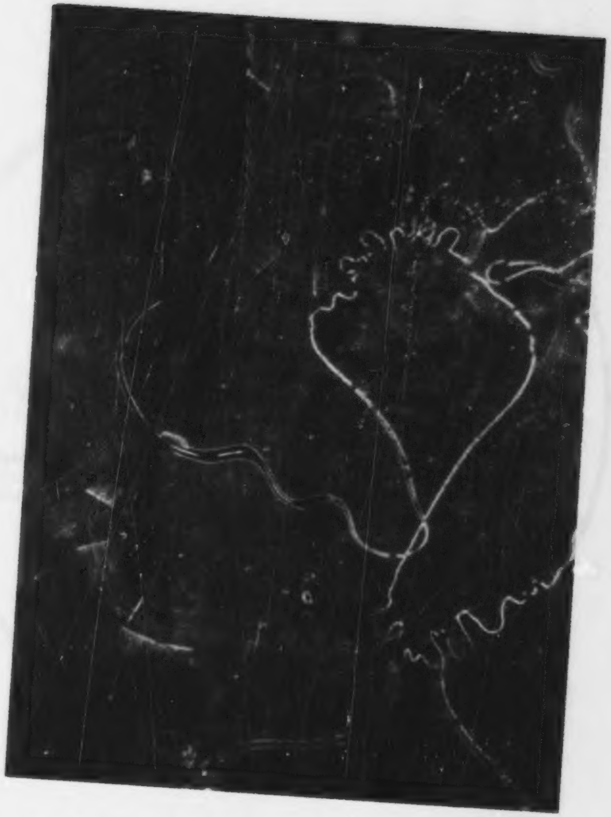


Figure 3 Droplet formation, filament breakup

For example, increasing conductivity from nominally 300 to 10,000 micromhos/cm requires a reduction in electrode voltage of 2-4 kv, but does not substantially change the characteristics of the spray. Apparently, the improved conductivity decreases the relaxation time for droplet charging.

Particulate collection in the CDS is accomplished by both direct collision of high energy water droplets and electrostatic precipitation. Direct collision and agglomeration is the dominate mechanism (efficiency) for particulate greater than one micron, depending upon particle density and wetability. Water droplets, instantaneously charged and accelerated, reach terminal velocities approaching 100 fps with a velocity trajectory which is substantially cross flow to the particulate velocity. The resulting high relative velocity (cross flow mass transfer) and optimally generated water droplet size (100-200 microns) result in target collecting efficiencies as high as 99+% for 2-3μ sized particle.

Two electrostatic particulate charging mechanisms are used to achieve submicron particulate collection. Charge transfer to the particulate is accomplished through gas ion current generated at the tube tips by corona discharge (conventional ESP) and by grazing encounters of the particulate with water droplets. In the latter case, charge transfer is the result of field distortions during the droplet-particle encounter which result in a droplet surface electrical breakdown. Such surface electrical breakdown can effectively occur at surface to particle distance of up to 2μ for a 0.1μ particle (e.g.: a single encounter will transfer sufficient charge to precipitate a particle at nominal operating conditions). The optimum relationship between particle and droplet size, Figure 4, has been estimated from theoretical considerations. The optimum droplet size is about proportional to the particle size, but always a factor of ten larger.

EQUIPMENT DESIGN

The previous section described the equipment configuration and operating parameters for a single chamber (module) and electrostatic spraying stage. CDS overall efficiencies are a function of process conditions and equipment design. Significant process conditions are particle size and composition. Physical properties of importance are specific gravity, wetability, and to a lesser extent, high electrical resistivity.

In considering overall equipment efficiency for a specified process the two most significant CDS design parameters are: Number of stages operated (Figure 5) and the plate spacing - gas flow velocity relationship, (Figure 6). Both of these parameters

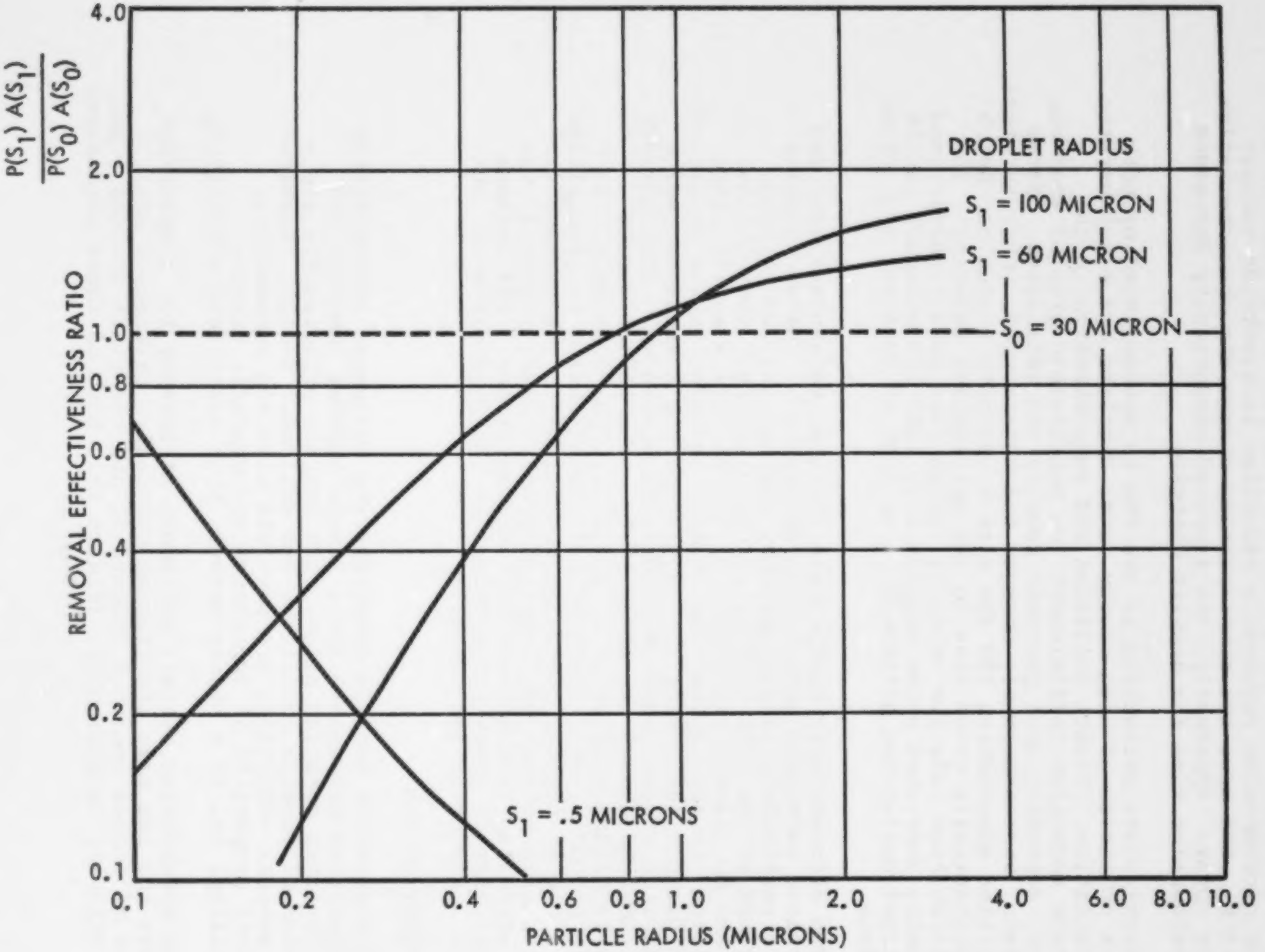


Figure 4 Theoretical Droplet-Particle Removal Effectiveness Relative to a 30 micron Radius Droplet

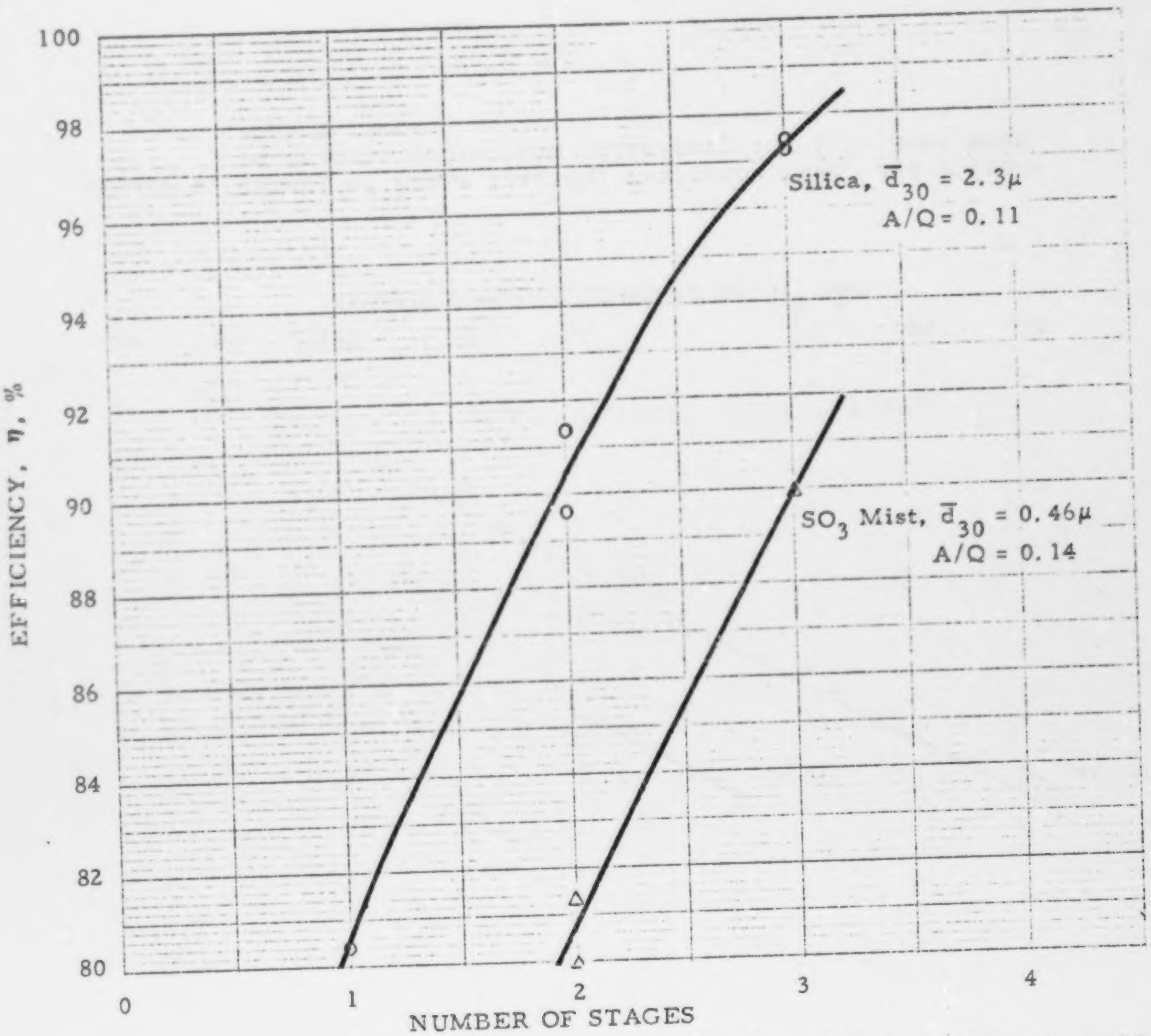


Figure 5 Effect of Series Electrostatic Spraying Stages on Overall Collecting Efficiencies

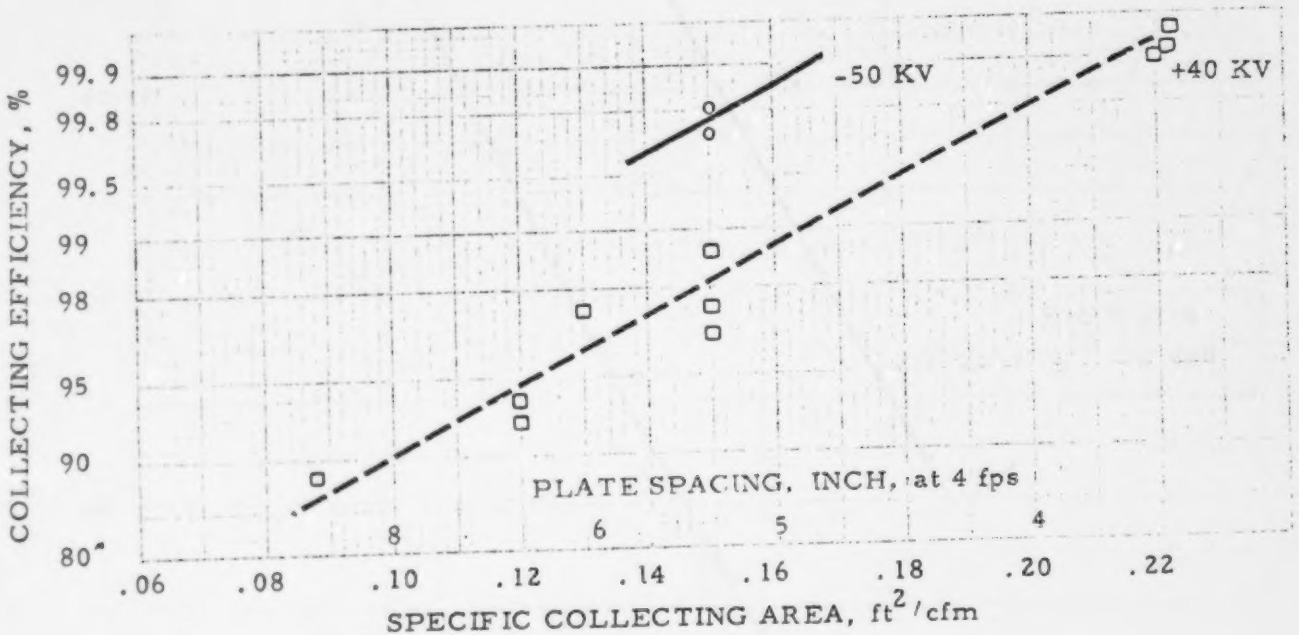


Figure 6 Effect of Plate Spacing, Residence Time, and Voltage on 3-Stage Collection Efficiency for 1.8μ Mass Mean Mineral Dust

are fundamentally bounded by the collector plate treatment height of eight feet, which was selected to meet the attractive commercial design requirements of preassembled, truck shippable structures and access for maintenance from a single work platform. Depending upon the characteristics of the process gas, which in turn affects the final plate spacings, this dimension allows either three or four electrical spray stages in series and, primarily, sets the baseline collecting efficiency. Greater efficiencies are achieved by concurrently selecting a plate spacing and gas velocity.

The collector plates can be arranged on 5 inch to 8 inch centers in increments of 1 inch to make the first efficiency adjustment. Operating voltages are set for spark rates of 100 to 250 arcs/min. Greater spark rate frequencies allow momentary field collapse and gas bypass. The possible benefits of higher voltage are consequently limited by the controlling spark rate criteria.

Gas flow capacity for a single module is governed by the plate spacing, the plate width and the gas velocity. Again considering the limitations imposed by contemplated truck transportation, the collector plate width has been fixed at 10 feet. This selected design dimension results in a nominal modular flow capacity of 2400 cfm at a base design plate spacing of 8 inches and a gas velocity of approximately 6 feet per second. All pilot CDS (1000 cfm nominal rating) incorporate the basic CDS electrical configurations consequently, there is no electrical scaling required to achieve full scale equipment design from pilot test data. In full scale equipment the flow capacity is controlled by preselected groups of modules arranged in parallel to conform to the standard equipment model sizes.

RECENT DEVELOPMENT AND OPERATING EXPERIENCE

The operating experience (pilot and commercial) with the first generation three stage CDS has been previously reported, References 1 and 3. This work identified a need for more economical equipment for independent and simultaneous fine particulate and noxious gases emissions control. Fortunately, these requirements need not be mutually exclusive (if the system is wet). Indeed, similar changes in equipment configuration and operation are required.

- Higher liquid-to-gas ratios (longer liquid surface residence times).
- Extended, wetted surfaces for absorption/space charge collection.

The first change to the baseline three stage design was to add a fourth stage within the same envelope. This is possible because the requirement for higher specific plates areas dictates closer plate spacing which, in turn, allows closer stage spacing without space charge interactions. A four stage commercial pilot is shown in Figure 7, with its auxiliary equipment.

Simulation tests with submicron iron oxide fumes substantiated improved collection efficiencies of the four stage unit.

The following pilot and commercial applications experience has subsequently been acquired on fine particle control.

- Pilot scale urea prilling tower.
- Commercial scale flue gases containing varying amounts of SO_2.

FINE PARTICLE CONTROL

Typically, the most difficult emission control requirement is to meet stack discharge opacities when substantial fractions of micron sized particles are present.

Urea prilling is a direct contact evaporative process for crystallizing urea droplets. Drying is accomplished by a long gravity free-fall between the injection point and the collection point. Rising drying air entrains sub-micron urea particulate and emits fume to the atmosphere. Aerodynamic mass mean particle sizes are in the range of .6 to 1.2 microns, (Figure 8). Dust loads are typically in the range of .012 and .038 gr/dscf, substantially out of compliance as opacities exceed 60%. The urea is very hydroscopic and soluble in water.

A four stage pilot scale TRW CDS was installed at a typical site for manufacturing urea by the prilling process. The pilot CDS was essentially identical in its internal design to a full scale, commercial unit, thus, no performance scale-up is necessary. Previous experience indicated that placing the collector plates on 6 inch centers would achieve the desired performance requirements. The performance was varied by adjusting the gas velocity between the collector plates. Breeching between the CDS and the process was accomplished through a length of about 150 feet and 12 inch pipe. No attempt was made to extract an isokinetic sample since the micron sized particulate should not stratify. The EPA Method 5 procedure was used to measure inlet and discharge particulate concentrations.

The test results have been generalized in a plot of

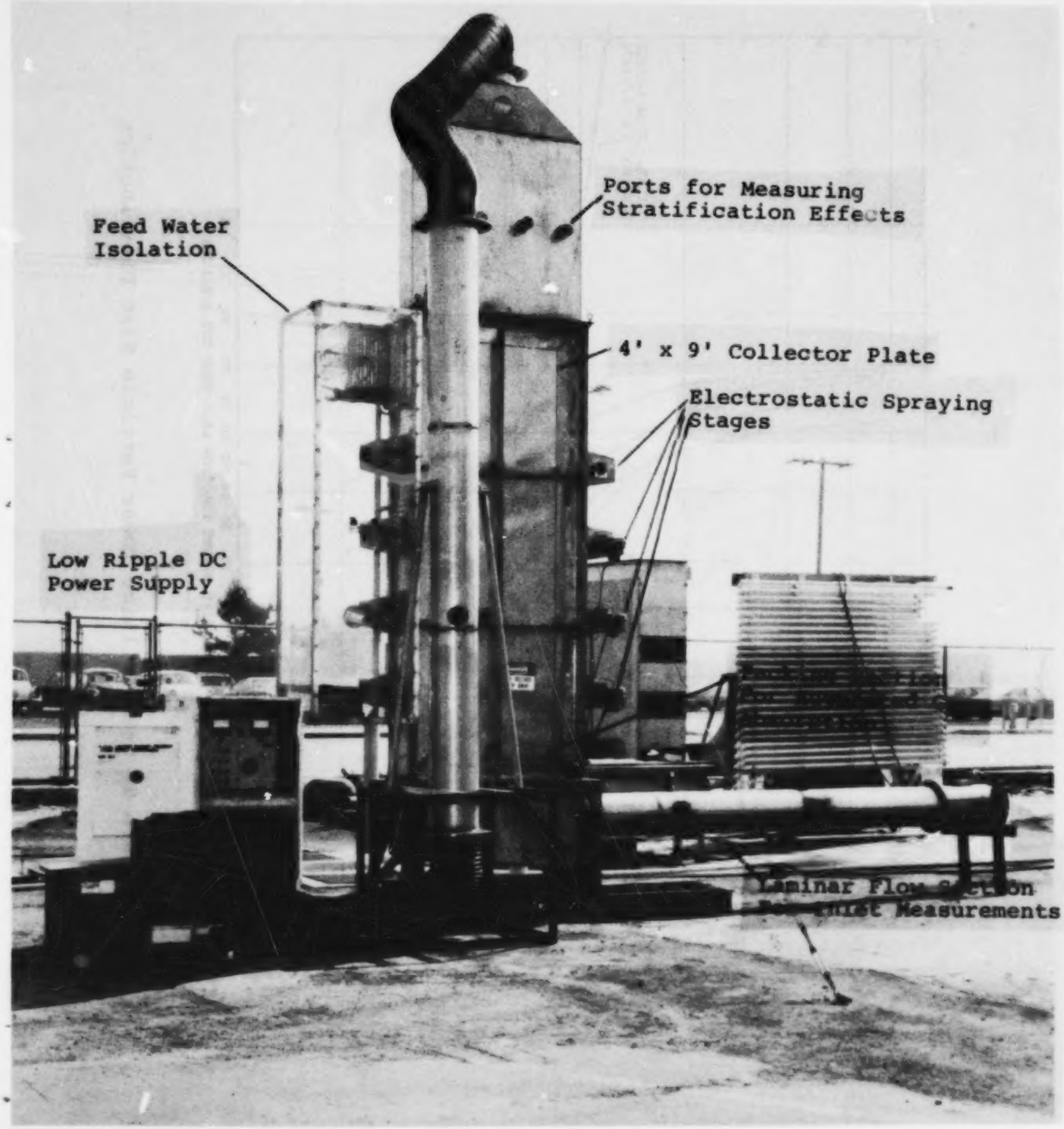

Ports for Measuring Stratification Effects
Feed Water Isolation
4' x 9' Collector Plate
Electrostatic Spraying Stages
Low Ripple DC Power Supply
Laminar Flow Section For Inlet Measurements

Figure 7 Pilot CDS, Full Scale Four-Stage Module Use for Fine Particulate and Gaseous Emissions Development Program

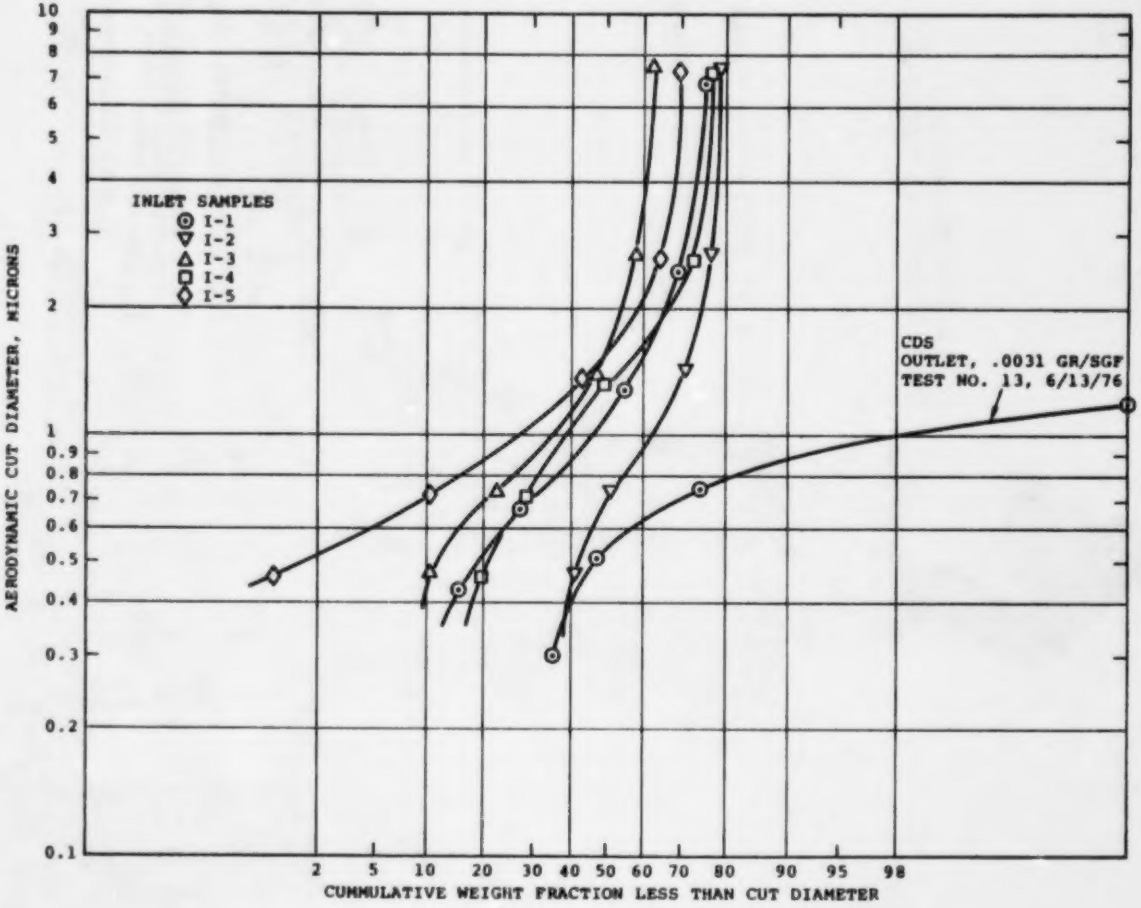


Figure 8 Urea Prilling Tower Particle Size Distribution

efficiency versus specific plate area (velocity), Figure 9. Efficiencies can be varied from 70% at a specific plate area and gas volocity of about .063 and 8.8 fps, respectively to 95% by derating velocity. The most cost-effective operating point on the curve is near its knee at about a .08 specific plate area. The pilot CDS stack discharge was substantially clear at this operating point. A downdraft configuration was also tested at the design efficiency to determine if performance improvements could be achieved. About a 2 to 3% efficiency increase at a lower electrode waterflow rate (about 25% reduction) was measured. The higher efficiencies are attributed to the fact that the lower water flow rates result in larger residence times of the water columns in the electrostatic field. This results in a smaller mean droplet size and improved collection efficiency of submicron particulate (see Figure 4). This configuration may also offer some significant economies for ground level installations by reducing the amount of duct-work required, particularly for an induced draft fan configuration. Higher electrode operating voltages (38 to 40 kv) improved efficiency at the design gas flow velocity. Utilities requirements at the design point were:

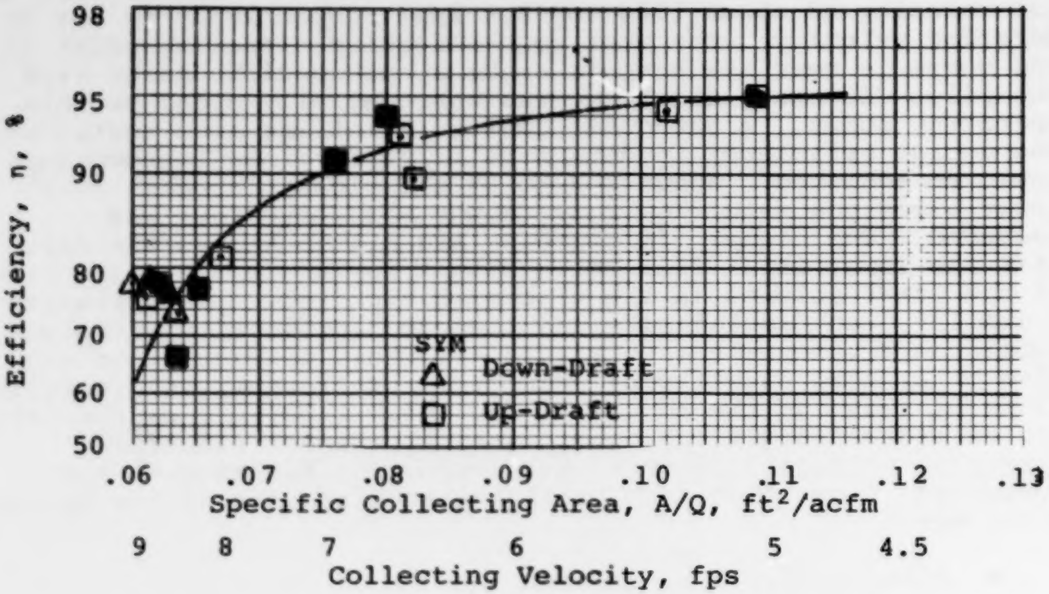
- Electrical, .4 watts/acfm
- Water, .46 gpm/1000 acfm

Process gas flange-to-flange pressure drops were nominally 0.5 inches wc and recycled water conductivity varied from 200 to 2400 micromhos/cm.

GASEOUS EMISSION CONTROL

The disadvantages of a wet system (disposal of a sludge) can be turned to advantage if simultaneous, selective absorption of noxious gases can be effected. A two-part development program has been completed to evaluate the feasibility of using the CDS to absorb the chemically active (HCl) and inactive (SO_2) gases. Chemically active systems require high mass transfer rates through the gas film boundary layer, inactive systems require improved mass transfer rates into the liquid surface. Typical process conditions are combustion flue gases containing 400 to 2000 ppm noxious gases such as SO_2.

For the reactive HCl system investigation a modified pilot scale CDS (four electrostatic spray stages and plates on 5 inch centers) was used. A hydrogen chloride high pressure cylinder was connected to a trace gas injector installed within a mixing chamber. The hydrogen chloride gas was distributed through three horizontal tube runs containing 15 equally spaced orifices. Sampling ports were located in the vicinity of the



PROCESS SPECIFICATION

Dust Load, gr/dscf	0.012 - 0.038
Temperature, °F	Ambient
Humidity, %	100
Particle Size - mean, μ	0.7 - 1.2
- % < 1μ	30 - 60
Particle Composition	Urea

CDS DESIGN AND OPERATION

Plate Spacing, inch	6
Gas Velocity, fps	4½ - 9
Transformer - Voltage, KV	50 - 55
- Power, va/cfm	0.4 - 0.8
Number of Stages	4

Figure 9 CDS Pilot Performance Urea Prilling Tower

inlet to the CDS where stratification of the injected gas was anticipated to be small. The CDS discharge sampling points were located at ground level just before the CDS induced draft fan.

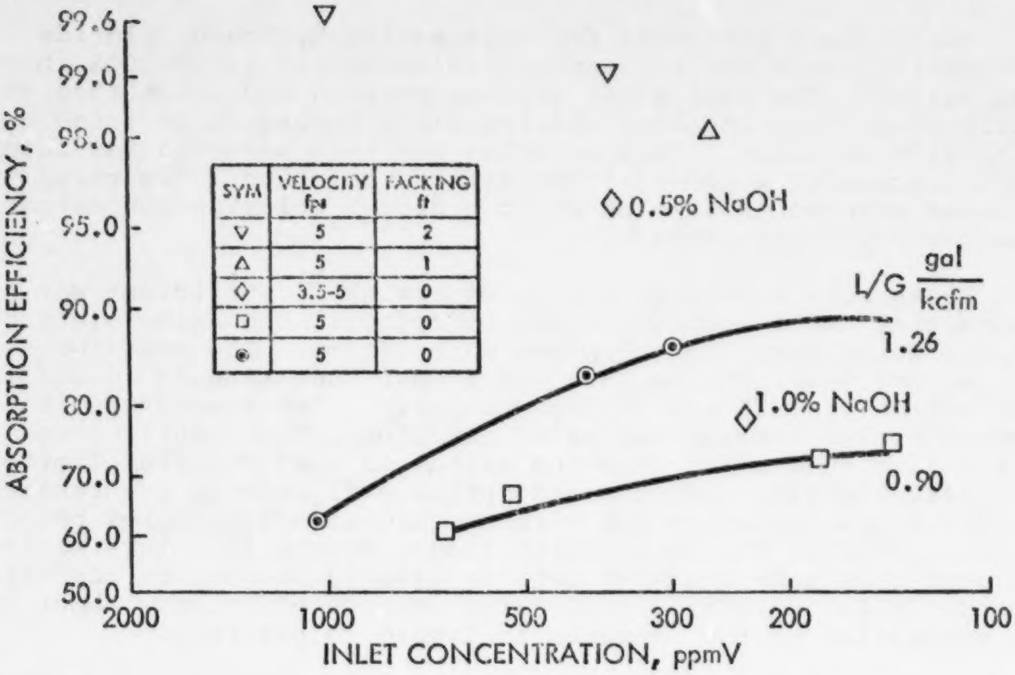
The primary procedure for determining hydrogen chloride absorption rates was to sample simultaneously at the CDS inlet and outlet. The sample was dry gas metered and drawn through a train consisting of three fritted glass impingers in a sodium hydroxide solution. This solution was then analyzed for chloride by titration of a known silver nitrate solution. The correlation between metered to sampled inlet hydrogen chloride concentrations was usually within ±20%.

Tests were initially run to determine if sufficient wet contacting and absorption could be accomplished using electrostatic water spraying and plate water films. The tests were run at two space velocities (3.5 and 5 fps) resulting in liquid to gas ratios of 1.26 and .9, respectively. The absorption is substantially removed (order of magnitude) from equilibrium saturation suggesting that the system is mass transfer limited. The data indicate that the absorption efficiencies are sensitive to inlet HCl concentration with the higher efficiencies being accomplished at the lower inlet loads, Figure 10. This would suggest that mass transfer rate is limited in part by the water droplet film. Higher liquid rates improve efficiency about 26%, in proportion to the increase in liquid to gas ratio.

To improve collection efficiencies by increasing the absorptivity of the liquid, a weak sodium hydroxide solution was applied through the plate wall wash system. The liquid to gas ratio for the plate wall wash is a factor of three higher than the electrode water so the comingled electrode spray water does not substantially dilute the concentration of sodium hydroxide. Total wetted surface area is not affected by this change so improvement in absorption must be attributed to improved liquid film mass transfer. This configuration allows recirculating the sodium hydroxide liquor through the wall wash system until a substantial fraction of the NaOH is consumed. Further efficiency improvements of up to 95% were achieved.

Interfacial CDS velocities are similar to those used in packed bed absorbers allowing a simple modification to affect a low pressure drop series equipment arrangement.

The current pilot and commercial configurations of the CDS have about 2 feet of open space between the gas distribution baffles and outlet flange within the lower section. This void space can be converted to a prescrubber by the addition of a non-plugging type of packing. In this case, the upper stage



PROCESS SPECIFICATION

HCl Concentrations, ppmV	180 - 1000
Temperature, °F	80 - 185
Humidity, %	20 - 30

CDS DESIGN AND OPERATION

Plate Spacing, inches	5
Gas Velocity, fps	3.5 - 5
Transformer - Voltage, kv	50
- Power, watt/cfm	0.9 - 1.2
Number of Stages	4

Figure 10 HCl Absorption Efficiencies for Several Equipment Configuration and Operating Variables

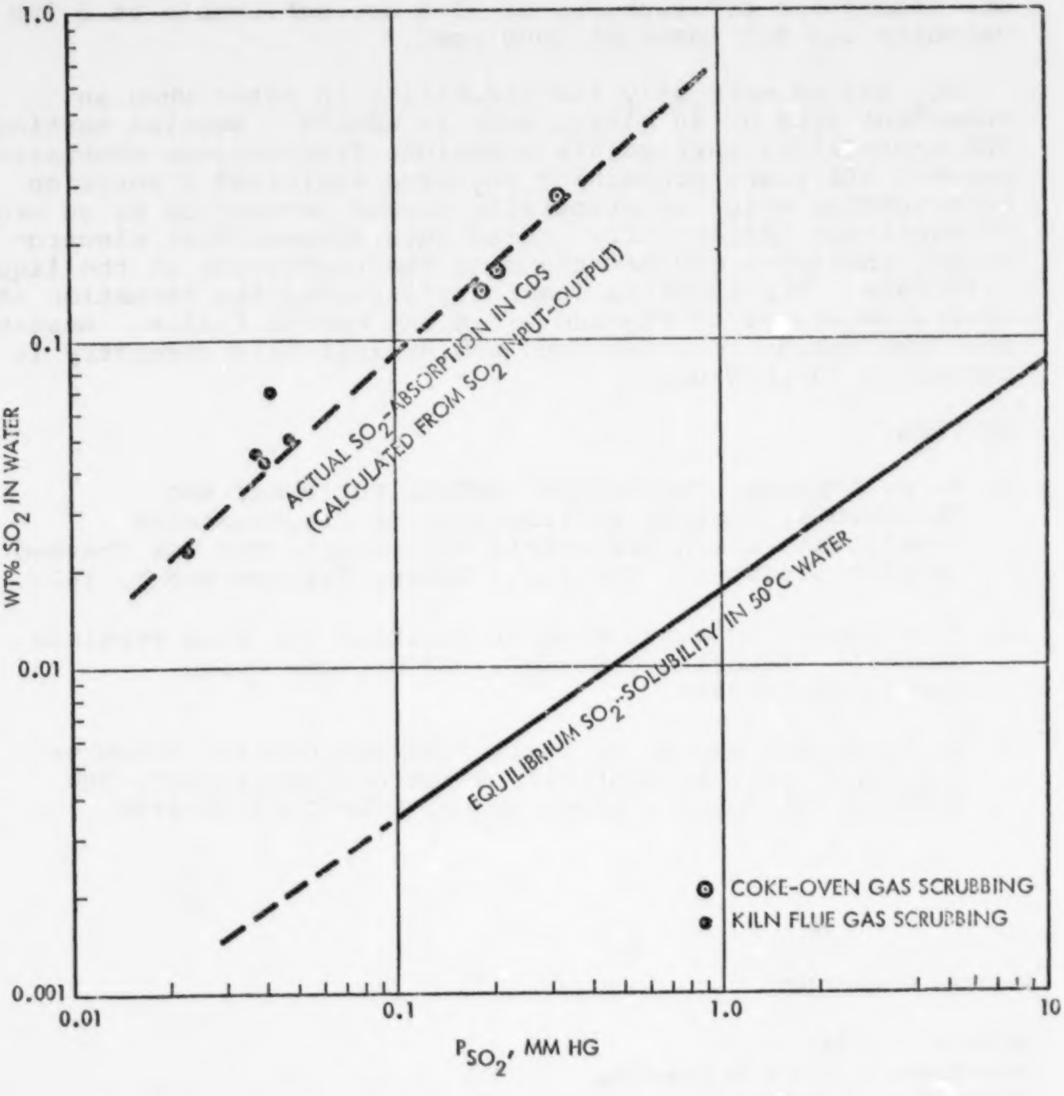


Figure 11 Preliminary Data Showing Unusual SO_2-Absorbing Capability of CDS

electrostatic spraying discharge water should be separated from the post-particulate scrubbing section to minimize the suspended solids clean-up for the recirculated sodium hydroxide solution. Liquid to gas ratios must be increased to about four, with the packed bed being operated at a L/G of 3. The total system pressure drop with the maximum packing depth of 2 feet was 1.5 inches wc. These test results, Figure 10, demonstrate that HCl absorption efficiencies of 99+% are achievable at 5 fps space velocity and HCl input of 1000 ppmV.

SO_2 has an extremely low solubility in water when an absorbant such as an alkali salt is absent. Routine testing of CDS controlling particulate emissions from various combustion process off gases containing SO_2 have indicated absorption efficiencies which substantially exceed saturation by an order of magnitude (Figure 11). These data suggest that electrostatic phenomena may be enhancing the reactivity at the liquid interface. Hypothesized chemistry includes the formation of activated states of SO_2 and O_2 in the corona fields. Research and development to understand and exploit this chemistry is currently in progress.

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1. R. R. Koppang, "POLLUTION CONTROL EQUIPMENT AND TECHNOLOGY, Broader Availability of Electrostatic Precipitators - A New Hybrid EP Concept, The TRW Charged Droplet Scrubber", TRW Inc., Energy Systems Group, 1975.

2. C. W. Lear, "Charged Droplet Scrubber for Fine Particle Control: Laboratory Study", TRW Systems Group, EPA-600/2-76-249a.

3. W. F. Krieve and J. M. Bell, "Charged Droplet Scrubber for Fine Particle Control: Pilot Demonstration", TRW Defense and Space Systems Group, EPA-600/2-76-249b.

WRITTEN DISCUSSION

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Department of Civil Engineering
University of Washington
Seattle, Washington

The paper presents the description and performance of a space charged scrubber-precipator named the "TRW Charged Droplet Scrubber." My comments concerning the paper are mainly in the form of questions; these questions being posed in order to better understand the scrubber system. My questions are as follows:

1. On page 3 a voltage gradient of 2.3 kV/cm is mentioned as necessary to achieve electrostatic spraying. Then on page 6 is it mentioned that increasing the water conductivity from 300 to 10,000 micromhos/cm requires a reduction in the electrical voltage of 2 to 4 kV. Why is this voltage reduction necessary?

2. In Fig. 4 what are the definitions of $P(S_1)$, $A(S_1)$, $P(S_o)$, and $A(S_o)$?

3. In Fig. 5 what are the units and definitions of A/Q?

4. The water to gas flowrate ratio of 0.46 gal/1000 acf seems somewhat low (page 14). Is this the water requirement for one stage, or is the water usage not including water recycling?

5. The electrical utility requirement is listed as 0.4 watts per acfm. Does this electrical usage include power required for water pumping, the fan, and the high-voltage power supplies?

6. Is the particle collection efficiency (either overall or at a specific particle size such as at 0.5 micron diameter) of the CDS sensitive to the inlet particle mass concentration? The collection efficiency data presented (Fig. 5, Fig. 6, and Fig. 9) only mentions the area prilling concentration in the 0.012 to 0.038 gr/scf.

For a comparison with the TRW CDS system, let us look at the U. of W. Electrostatic Scrubber (UWES). A schematic illustration of the UWES is provided in Fig. 1. The UW Electrostatic Scrubber involves the use of electrostatically charged water droplets to collect air pollutant particles electrostatically charged to a polarity opposite from the droplets. The particles are electrostatically charged (negative polarity) in the corona section. From the crona section the gases and charged particles flow into a scrubber chamber into which electrostatically charged water droplets (positive polarity) are sprayed. The gases and some entrained water droplets flow out of the spray chamber into a most eliminator consisting of a positively charged corona section in which the positively charged water droplets are removed from the gaseous stream.

The general layout of the UW Electrostatic Scrubber pilot plant (Mark 2P model) is shown in Fig. 2. The system (in the direction of gas flow) includes a gas cooling tower, an inlet test duct with sampling port, a particle charging corona section (corona no. 1), a charged water spray tower (tower no. 1), a particle charging corona section (corona no. 2), a charged water spray tower (tower no. 2), a positively charged corona section to collect the positively charged water droplets, an outlet test duct with sampling port, and a fan. The pilot plant is housed in a 40 foot long trailer and can be easily transported to emission sources.

In Fig. 3 the particle collection efficiency as a function of particle size

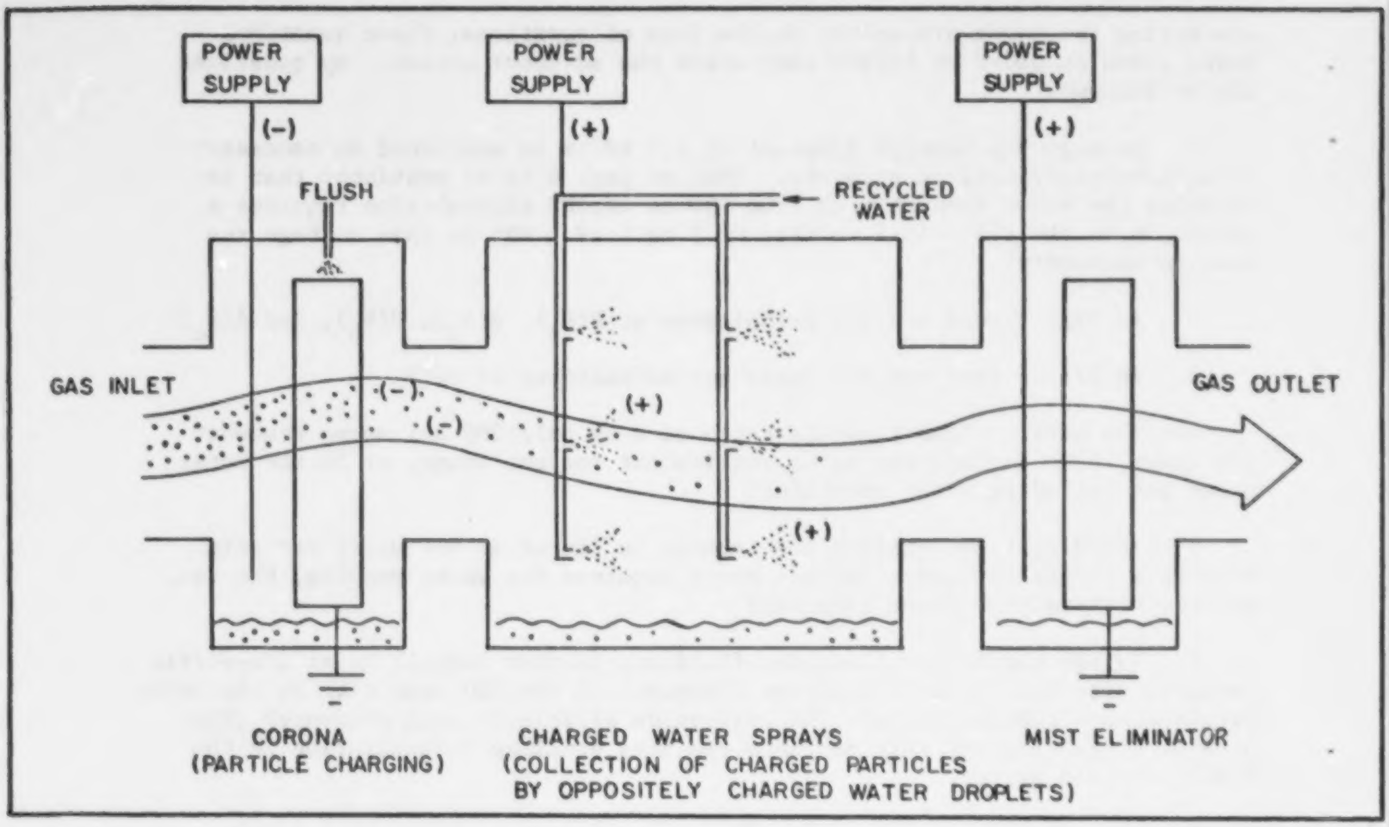

POWER SUPPLY
(-)
POWER SUPPLY
(+)
POWER SUPPLY
(+)
FLUSH
RECYCLED WATER
GAS INLET
(-)
(-)
(+)
(+)
GAS OUTLET
CORONA
(PARTICLE CHARGING)
CHARGED WATER SPRAYS
(COLLECTION OF CHARGED PARTICLES
BY OPPOSITELY CHARGED WATER DROPLETS)
MIST ELIMINATOR

Fig. (1) UW Electrostatic Scrubber

is presented for emissions from an electric arc steel furnace. Note that the specific collecting area of the UWES is .05 ft^2/acfm which is somewhat less than the .08 to .22 ft^2/acfm presented by Koppang for the CDS system. Of course this illustrates the difference in the particle collection operating mechanisms utilized, the UWES using charged water droplets to collect the particles and the CDS having a greater collection plate area. Fig. 4 and Fig. 5 present data obtained at a pulverized coal-fired power plant. Fig. 4 illustrates the increased collection efficiency obtained at the higher water charging voltage. Fig. 5 presents the effect of water to gas flowrate ratio on the collection efficiency.

For additional information on the UWES system, see the references listed below:

[1] Pilat, M.J., S.A. Jaasund, and L.E. Sparks (1974) "Collection of Aerosol Particles by Electrostatic Droplet Spray Scrubbers" Envir. Sci. & Tech. 8 340-348.

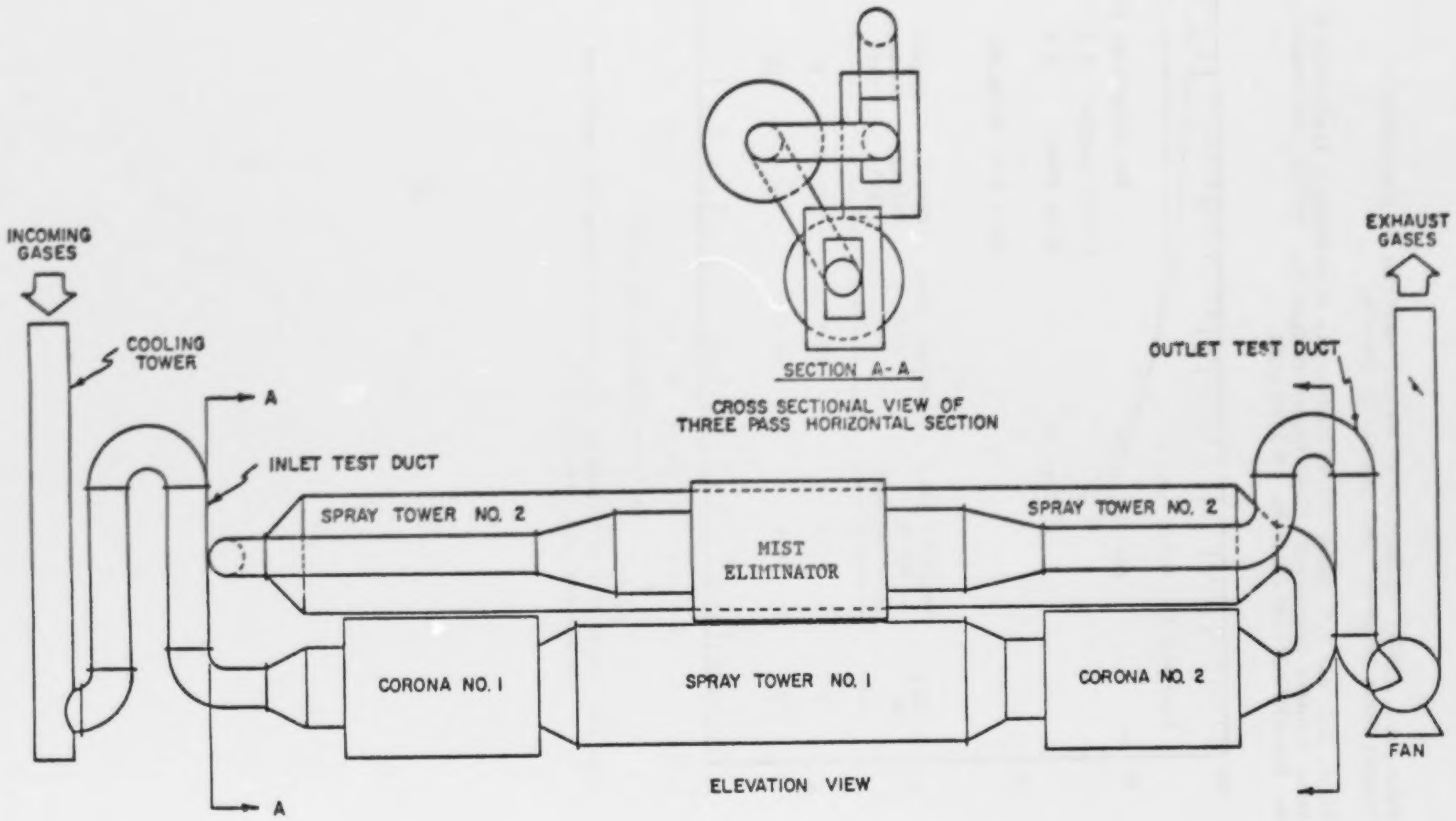
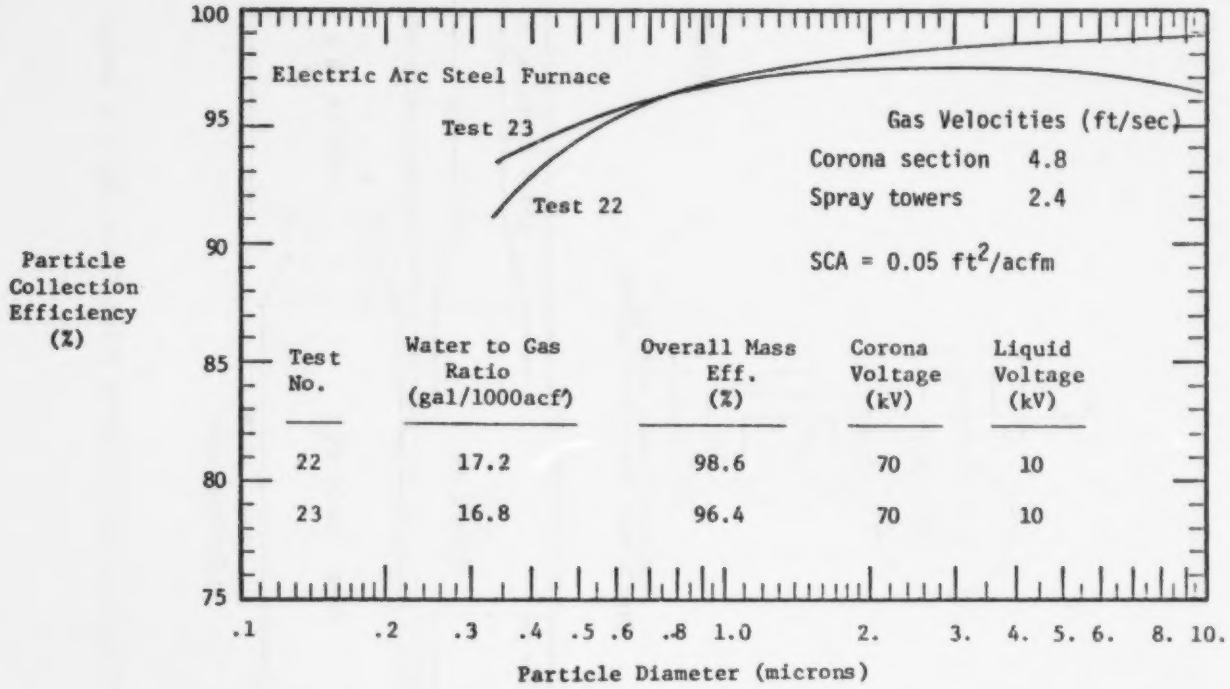

SECTION A-A
CROSS SECTIONAL VIEW OF THREE PASS HORIZONTAL SECTION
INCOMING GASES
COOLING TOWER
A
INLET TEST DUCT
SPRAY TOWER NO. 2
MIST ELIMINATOR
SPRAY TOWER NO. 2
OUTLET TEST DUCT
EXHAUST GASES
CORONA NO. 1
SPRAY TOWER NO. 1
CORONA NO. 2
FAN
A
ELEVATION VIEW

Fig. 2 General Layout of Electrostatic Scrubber Pilot Plant (Mark 2P Model)

[2] Pilat, M.J. (1975) "Collection of Aerosol Particles by Electrostatic Droplet Spray Scrubber" APCA Journal 25 176-178.

[3] Pilat, M.J. and D.F. Meyer (1976) "University of Washington Electrostatic Spray Scrubber Evaluation" Final Report on Grant No. R803278, EPA Report No. EPA-600/2-76/100 (NTIS No. PB 252653/AS).



Test No.	Water to Gas Ratio (gal/1000acf)	Overall Mass Eff. (%)	Corona Voltage (kV)	Liquid Voltage (kV)
22	17.2	98.6	70	10
23	16.8	96.4	70	10

Fig. (3) Collection Efficiencies With Similar Operating Conditions At Electric Arc Steel Furnace

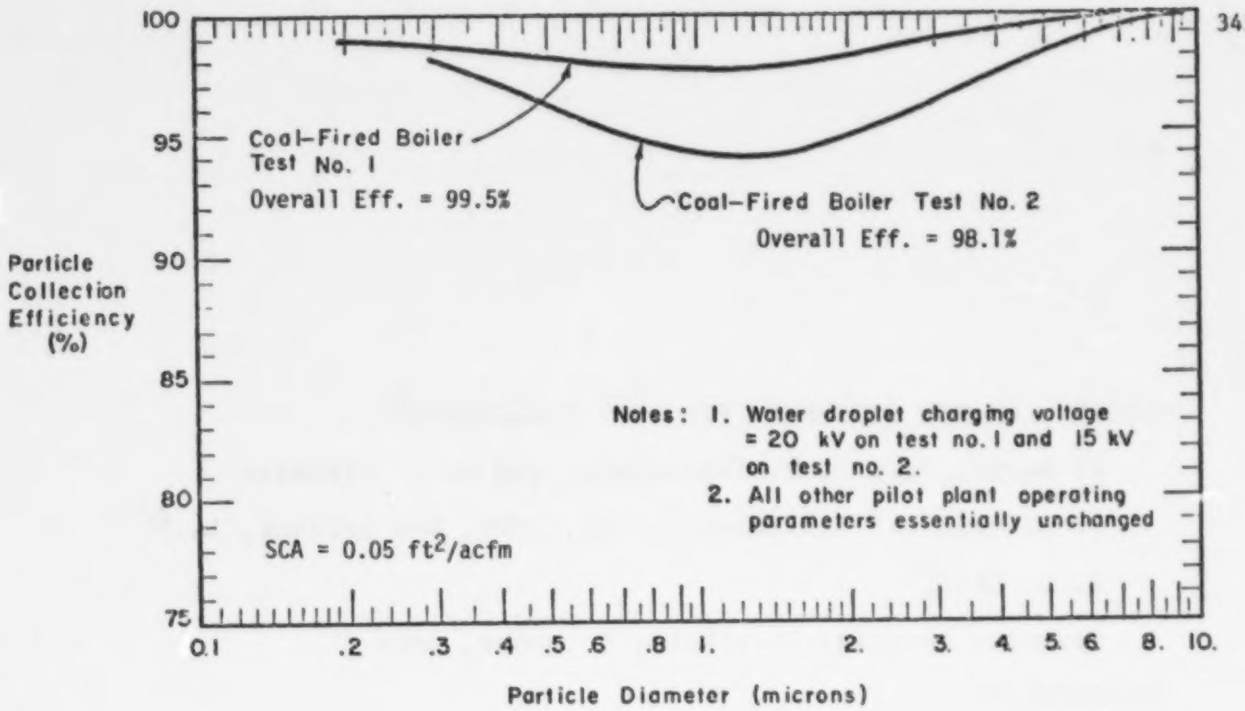


Fig. 4 Influence of Water Droplet Charging Voltage on Particle Collection Efficiency

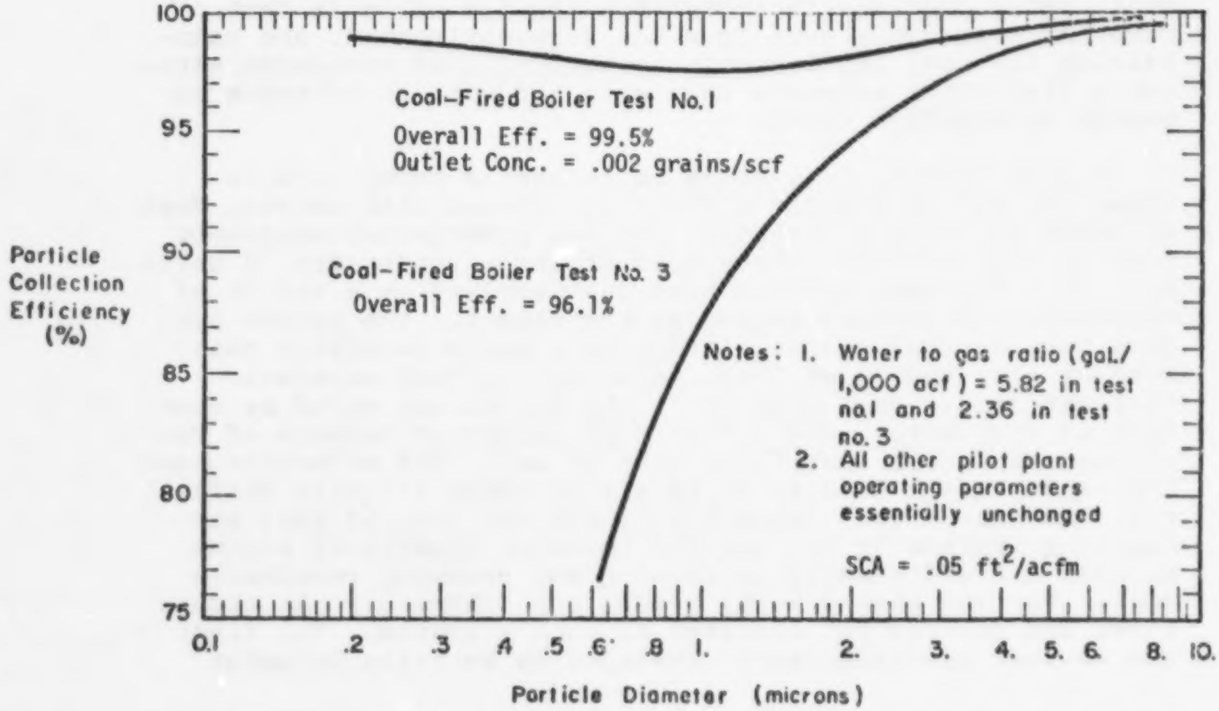


Fig. 5 Influence of Water to Gas Flow Rate Ratio on Particle Collection Efficiency

RESEARCH ON NEW EQUIPMENT FOR DUST COLLECTION[1]

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INTRODUCTION

OSHA cotton dust standards and the possible occurrence of byssinosis in cotton textile workers has focused attention on the dust problem in cotton textile mills and related operations where cotton and cottonseed are processed. One of the most urgent problems facing cotton textile mills is that of removing respirable dust from the work environment and maintaining the dust level within prevailing OSHA standards without a disturbing economic unbalance and with no increase in energy consumption (1-4).

A wide variety of systems is currently being used to clean the air in textile mills that process lint cotton. Most of these are custom designed, and use proprietory equipment such as bag filters, rotary drum filters, condensers, V cells etc. (5), but many systems have just evolved as a result of additions and changes needed at the moment. The better systems have several stages, including a waste separator that removes lint and large trash particles, a dust separator that can reduce dust content of the air to one mg/m^3 by some form of filtration, and a fine dust separator capable of removing respirable dust (less than 15 µm). The effective control and removal of dust in this size range is quite difficult and expensive. Presently, there are few, if any, air handling systems in the textile industry capable of economically and efficiently collecting and removing respirable dust. For the past several years, our laboratory has sponsored and carried out research to such a system. The first two devices described were developed by Battelle Columbus

Laboratories under contract to U.S.D.A., and the third one was developed at the Southern Regional Research Center in New Orleans.

Wet Wall Electroinertial Unit

The first device, called a wet wall electroinertial unit (WWEU) air cleaner, was developed to meet the requirements of high operating efficiency, low maintenance, low operating cost, and simplicity of design. To achieve high efficiency with low pressure loss, an electrostatic precipitation system was the best candidate. A wet wall was desirable to flush the precipitated material away and to minimize maintenance. Inertial effects were added to assist in moving material toward the wall and to improve the cleaning action of the water on the wall.

These features were best incorporated in the design shown in Figure 1. The unit consists of a concentric wire-in-tube precipitator. The charging wire is located axially in a stainless steel tube in a vertical orientation. Air enters tangentially at the upper end of the tube (through a duct with a flattened cross section) and acquires a rotating movement. When high voltage is applied to the wire, an ionizing corona discharge charges the dust in the air. As the air flows through the tube, the charged dust is driven to the wall by the radial electrical field. Water from the upper water inlet flows down the wall and flushes the precipitated dust into the water outlet at the lower end of the tube. The rotational movement of the air also induces rotational flow of the water, which assists in uniformly wetting the surface of the tube. Cleaned air is expelled from the tube's lower end.

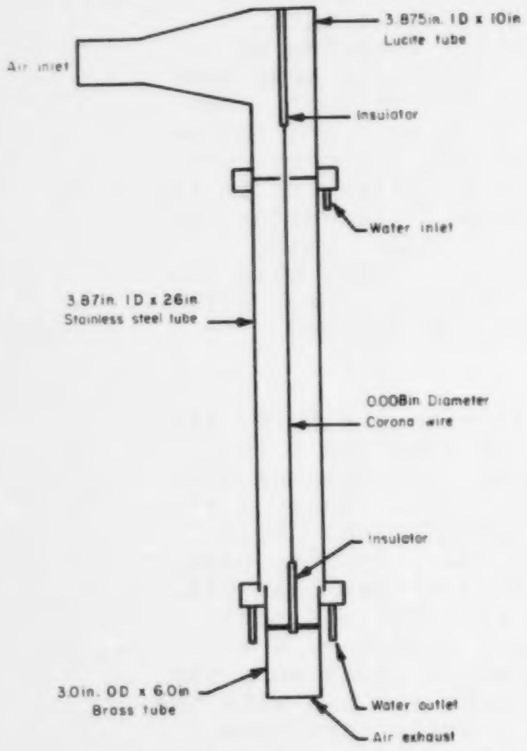
3.875in. ID x 10in
Lucite tube
Air inlet
Insulator
Water inlet
3.87in. ID x 26in
Stainless steel tube
0.008in Diameter
Corona wire
Insulator
3.0in. OD x 6.0in
Brass tube
Water outlet
Air exhaust

Fig. 1. Wet Wall Electroinertial Unit

Tests of the WWEU were conducted with 4 in. and 8 in. diameter units at flow rates ranging from 100 to 800 ft^3/min. Particle collection efficiency of these units was measured with atmospheric dust, AC fine test dust[4] (mass mean diameter (MMD) = 12 μm), artificial cotton dust (MMD = 4.0 μm), and cotton dust drawn from the processing area of a card (MMD = 3.0 μm). Efficiency data with atmospheric dust was obtained by isokinetic-sampling downstream of the unit with an optical particle counter. Sample preparation and testing procedures with the AC test dust and artificial cotton dust have been described previously (6). Essentially, known quantities of these dusts were fed to the WWEU and all particles that escaped were captured on an absolute filter installed downstream of the unit. Preliminary results are also included on tests with an 8 in. WWEU that cleaned dust laden air drawn from a full scale cotton card operating at 20 lb/hr.

Tests were conducted with the 4 in. diameter WWEU to determine the effects of tube length, air flow rate, and voltage polarity on efficiency of particle collection. Tube lengths of 1, 2, and 4 ft were tried. Shearing of water droplets from the wall at higher velocities made the practical upper limit on flow rate about 200 ft^3/min. Results of the efficiency tests with the 2 ft unit are included in Table I. Runs made at zero potential and zero current with the rather coarse AC dust show efficiencies in the 80% range due merely to centrifugal forces. Increasing the voltage to -30 kV raises the efficiency to 99%. Which diminishes to 98.5% when +33 kV is applied. Tests with the fine artificial cotton dust indicate a somewhat lower efficiency (98.1 and 96.1% respectively) for 100 and 200 ft^3/min.

Table II contains the data for tests with the 8 in. diameter, 66 in. long WWEU. These data show that the 8 in. unit can be scaled up from the 4 in. WWEU, maintaining the same efficiency levels for a higher flow rate. While scaling up the capacity of the units, the velocity and field strength were maintained at the same values. In this case, particle collection efficiencies are still better than 95%. Differences in efficiencies with the artificial and real cotton dust for zero voltage are no doubt caused by the relative fineness of real cotton dust, which diminishes the effects of centrifugal forces. More recent studies with the 8 in. unit to clean air from an actual textile card show that the unit operates quite satisfactorily under realistic test conditions.

TABLE I

TOTAL MASS EFFICIENCY OF THE 4 IN. WET WALL ELECTROINERTIAL UNIT a/ b/

Test Dust	Circulation Rate ft^3/min	Potential kV	Current mA	Dust Feed Rate g/min	Pressure Drop in. H_2O	Efficiency %
AC fine	100	0	0	1.0	1.25	81.2
AC fine	200	0	0	2.0	3.75	87.6
AC fine	100	-30	2.4	0.2	1.25	99.0
AC fine	200	-30	2.4	0.2	1.25	99.0
AC fine	100	+33	--	1.0	1.25	98.5
Cotton dust	100	-30	2.2	0.83	1.25	98.1
Card trash	200	-30	2.2	0.67	3.75	99.7
Cotton dust	200	-30	2.1	0.62	3.75	96.1

a/ Length of WWEU was 2 ft.
b/ Water flow was 0.25 gal/min.

TABLE II

TOTAL MASS EFFICIENCY OF 8 IN. WET WALL ELECTROINERTIAL UNIT a/

Test Dust	Potential kV	Current ma	Water Flow g/min	Air Flow ft^3/min	Efficiency %
AC fine dust	-60	--	0.46	800	97.6
Card trash	-60	-7	0.46	800	98.4
Card trash	0	0	0.46	800	99.8
Artificial cotton dust	-60	-7	0.46	800	99.4
Artificial cotton dust	0	0	0.46	800	80.2
Card output b/	0	0	0.64	800	30.3
Card output b/	-58	-6	0.65	800	99.9
Card output b/	+60	+5.8	0.65	800	96.9

a/ Length of WWEU was 66 in.
b/ Tested at SRRC on output of V-cell filter.

High Speed Rotary Drum Air Cleaner

The air cleaning systems of most textile mills utilize a rotary drum (low speed) filter that allows the unit to continuously self-strip the accumulated filter cake. This procedure allows for a low pressure drop in the air circulating system but leads to inefficient dust removal, particularly at the respirable dust range. Investigation of these cleaners with the intention of increasing their efficiency led to application of the principle that movement of a collection surface normal to the direction of the air flow through it increases the effective interception cross section of the collector surface. If the collector surface moves at a velocity n times that of the air stream, the number of particles impacted on the surface is increased by a factor of between n and n + 1.

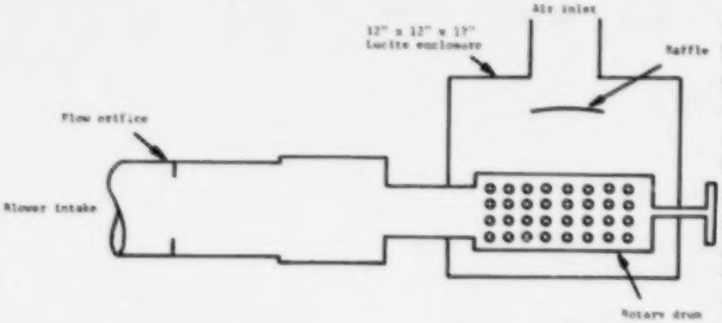

Fig. 2. High Speed Rotary Drum Air Cleaner

The apparatus shown in Figure 2 was constructed to evaluate this concept. A hollow aluminum cylinder was mounted coaxially on a drive shaft on one end and was connected via a large bearing to a 3 in. diameter suction pipe on the other end. The surface of the aluminum cylinder was perforated with holes arranged in equally spaced rows parallel to the axis of the cylinder. Filter materials were wrapped around the drum to cover the holes, and the drum was rotated to increase the effective cross section of the fibers. In operation, air was drawn in through the holes and out through the 3 in. diameter suction pipe by a high pressure blower as the cylinder was rotated. The efficiency of the rotary drum unit was evaluated for several configurations, such as hole size, open area, rotational speed, flow rate, and cylinder diameter. The drums were evaluated with and without filter media on the surface, and some of the drums were evaluated with electrostatic assist.

Initially, the efficiency of the rotary drum unit was evaluated by using a particle monitor to measure the particulate concentration in the air in the room outside the drum and in the air at a point inside the drum. During this work, some variation was observed depending on the position of the sampling probe inside the drum. These data generally reflect relative dust loading differences in the portion of the air flow sampled and do not indicate overall efficiency for the total air flow. The true efficiency of the unit was measured with AC fine test dust with a fiberglass filter installed downstream of the rotary drum to collect all of the dust that passed through it.

The first drum tested was a 4.5 in. OD x 10 in. long cylinder with a 5/32 in. wall. Sixty 1 in. diameter holes were drilled in ten equally spaced rows, with six equally spaced holes in each row to provide a total open area of 0.327 ft^2. The partial efficiency was remarkably high with the sampling probe inside the drum at a position on the axis of the drum and 3 in. from the bearing. As shown in Table III, from 58% to 93% of the atmospheric dust was removed from the air passing through this area when the drum was rotated at a surface speed of 3000 ft/min with air flow rate of 65 ft^3/min. The efficiency increased with drum speed up to about 3,000 ft/min; no further improvement was attained at 4,000 ft/min. The efficiency decreased as the air flow rate was increased to 200 and 300 ft^3/min. Particle counts outside the drum showed increased dust concentration in the air around the drum.

With a layer of polyester filter material covering the drum, the partial efficiency of the system was generally higher than that of the plain drum in the smaller particle size ranges. The efficiency in the larger particle size ranges is quite good (>90%), but is not significantly better than for the plain drum. Covering the drum with a PC (Pneumafil Corp.) second stage filter gave almost the same efficiencies as the aplin drum, whereas covering with a PC first stage filter resulted in a decrease in efficiency of the plain drum.

Two types of electrostatic systems were evaluated in conjunction with the rotary drum with 1 in. holes. In the first system, electrical fields were applied to the drum from a 1/4 in. hardware cloth cylindrical grid, spaced 1 in. from the drum. With potentials up to 20 kV on the grid, the efficiency of particle removal with the plain

TABLE III

PARTIAL EFFICIENCY OF ROTARY DRUM WITH 1-IN. DIAMETER HOLES

Drum Speed ft/min	Flow Rate, ft^3/min	Cover	Efficiency, percent $>$0.5 micron	$>$5.0 microns
1,000	65	None	4.1	23.5
2,000	65	None	42.3	85.9
3,000	65	None	58.2	93.1
4,000	65	None	51.4	93.5
1,000	200	None	3.1	12.7
2,000	200	None	9.3	55.4
3,000	200	None	16.1	60.8
0	65	E2B-P01B	16.9	49.1
1,000	65	E2B-P01B	63.8	95.4
2,000	65	E2B-P01B	75.7	95.4
3,000	65	E2B-P01B	82.6	94.6
0	65	PC, first stage	1.1	3.9
1,000	65	PC, first stage	29.3	81.4
2,000	65	PC, first stage	41.6	83.0
3,000	65	PC, first stage	51.8	86.9
0	65	PC, second stage	3.0	32.7
1,000	65	PC, second stage	48.9	90.6
2,000	65	PC, second stage	59.3	93.5
3,000	65	PC, second stage	61.1	93.4

drum increased to the level achieved with filters covering the drum. With the external field and filters on the drum, the increase in removal of submicron particles was about 10% to 20%.

In the second electrostatic system, a series of 0.005 in. diameter charging wires were spaced at about 2 in. intervals parallel to and 1 in. from the drum. Data in Table IV show that the partial efficiency was higher with auxiliary electrostatic charging. With the drum stationary, the charging system alone removed 56% to 73% of the atmospheric dust at 65 ft^3/min; however, when the flow rate was increased to 200 the efficiency decreased although not as much as without the electrostatic assist.

At low flow rates, the good performance of the E2B-P01B filter was improved considerably by adding the electrostatic assist.

TABLE IV

PARTIAL EFFICIENCY OF ROTARY DRUM (1-IN.-DIAMETER HOLES WITH AUXILIARY ELECTROSTATIC SYSTEM)

				Efficiency	
Drum Speed ft/min	Flow Rate ft3/min	Cover	Potential kV	> 0.5 micron %	> 5.0 microns %
0	65	None	-20	55.9	73.0
0	200	None	-20	41.7	52.8
1,000	65	None	-20	62.7	82.7
2,000	65	None	-20	76.7	90.1
3,000	65	None	-20	85.4	96.1
4,000	65	None	-20	75.6	95.1
1,000	200	None	-20	45.2	66.6
2,000	200	None	-20	49.2	79.6
3,000	200	None	-20	49.0	77.3
0	65	E2B-P01B	-10	14.5	20.6
0	65	E2B-P01B	-15	40.4	46.9
1,000	65	E2B-P01B	-15	84.9	96.3
2,000	65	E2B-P01B	-15	89.9	96.3
3,000	65	E2B-P01B	-15	92.3	96.4

The second drum tested (45 in. OD x 10 in.long with 5/32 in. wall) had 897 holes 1/4 in diameter drilled in 39 equally spaced rows with 23 equally spaced holes in each row to provide a total open area of 0.305 ft^2. This drum was considerably more efficient than the drum with 1 in. holes for removing atmospheric dust at lower speeds.

A third cylinder (same dimensions in the first 2) was made with additional 1 in. diameter holes to provide more open air: 91 holes were arranged in 13 rows of 7 holes each to provide an open area of 7 holes each to provide an open area of 0.496 ft^2. A flow rate of 65 ft3/min produced a gas velocity of about 130 ft/min through the holes in this drum, as compared with velocities of 200 to 210 ft/min with the first two. In general, performance of the drum with 91 1-in. diameter holes was erratic. Efficiency usually improved when surface speed was increased and worsened when air flow was increased. However, the overall performance was inferior to that of the drums with fewer or smaller holes.

Tests with AC fine dust were run with the drum enclosed in a 12 x 12 by 17-in. Lucite box with a 4 in. diameter air inlet in the center of the top of the box, as shown in Figure 2. A 4 in. diameter spherical deflector disk was mounted about 6 in. below the air inlet to distribute the incoming air around the box. AC fine dust was fed into a 4 in. diameter pipe connected to the inlet of the box by means of a small rotary platform feeder. Efficiency was determined by collecting material that passed through the drum on a 1/2 in. PF 105 Fiberglas filter downstream of the drum. Data from these tests run on the drum with both the 1 in. and the 1/4 in. holes is shown in Table V. For the drum with the 1 in. holes, the mass efficiencies agree quite well with the partial efficiencies for atmospheric dust. The mass efficiencies for the drum with the 1/4 in. holes do not agree with the partial efficiencies obtained with the optical counter since there is no evidence of a maximum at the low flow rates.

TABLE V

MASS EFFICIENCY OF THE ROTARY DRUM WITH AC DUST

Drum Speed ft/min	Hole Diam., in.	Flow Rate, ft^3/min	Cover	Potential, kV	Efficiency %
3,000	1.0	65	None	0	90.0
3,000	1.0	65	None	-20	93.0
3,000	1.0	65	PO-1B	0	98.0
325	0.25	65	None	0	57
805	0.25	65	None	0	72.3
1,530	0.25	65	None	0	80.8
3,000	0.25	65	None	0	90.3

The various measurements and observations made while working with the rotary drum indicate that separation of particulates occurs externally and internally. Some particulates are rejected from the air stream at the surface of the drum;

as a result, the particulate concentration in the air around the drum increases. At the same time, particulate concentration inside the drum varies at different points. Lowest particle concentrations generally occur along the axial center of the drum. At first the air stream was sampled along the axial center of the drum to avoid particulates generated by the bearing or introduced by air leaks through the unsealed bearing on the exhaust end of the unit. However, as work progressed, evidence indicated that particulate concentration in the air inside the drum was higher in off-axis positions. A functional design thus must provide means for removing the dust-laden air outside the drum and for separating the clean air from the dust-laden air inside the drum.

Fluid Electrode Precipitator

In general, the best way to remove fine dust from an air stream is to use conventional electrostatic precipitators. Although the efficiencies of these devices are quite high, there are inherent drawbacks that do not make them acceptable in continuously operating air systems. One of the major drawbacks to self-cleaning electrostatic precipitators is that they must be shut down in order to clean themselves. The development of the fluid electrode precipitator is an attempt to alleviate this problem. The fluid electrode precipitator illustrated in Figure 3 consists of an array of electrodes in a casing to charge, direct, and collect fine dust. Tungsten wires serve as discharge electrodes for all units tested, auxiliary driving electrodes are charged metal plates or screens that can be inserted between the fluid electrodes, and the collection electrodes are formed by vertical tubes that release a laminar flow of water falling through the height of the casing.

The precipitator operation is quite simple. The falling columns of grounded water act as cylinders in cross flow and create vorticies for enhanced particle collection. Auxiliary electrodes and discharge electrodes can be positioned in arrays to direct charged dust into the grounded fluid.

The use of free-falling fluid results in a system that is less complex than rigid surfaces with wetting devices. Shear forces due to air flow over a wetted surface can severely disrupt the thin layer of fluid flow. A cylindrical fluid column is a stable shape in cross flow, and is deflected very little by drag forces at velocities up to 1000 ft/min. The fluid is deflected by electrostatic attraction, but this can be compensated for by auxiliary electrode placement.

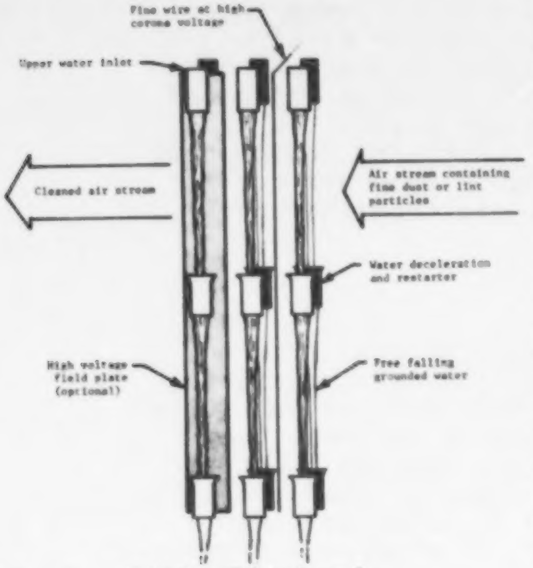
Fine wire at high corona voltage
Upper water inlet
Cleaned air stream
Air stream containing fine dust or lint particles
Water deceleration and restarter
High voltage field plate (optional)
Free falling grounded water

Fig. 3. Fluid Electrode Precipitator

The fluid electrode precipitator concept was applied to experimental model units, and collection efficiency of precipitator array configurations was measured in a cotton dust test facility. This apparatus consists of a 9 x 9 in. straight tunnel 9 ft. long, containing traversable sampling probes located upstream and downstream of the precipitator test section. Flow straighteners near these locations allow isokinetic sampling through probe tips of various inlet sizes. Dust analysis based on projected area diameter was performed with an optical particle counter. A mass monitor was used to analyze dust mass concentration.

Two series of tests, one at low capacity and one at high, were performed with cotton dust from different sources. For low capacity tests, air was drawn through a model-size cotton carding machine running at 2 lb/hr. The air was prefiltered to remove lint and large particles and was then passed through the test tunnel. In the high capacity tests, fine dust released by mechanically tapping a V-cell nonwoven filter was used. The filter was already loaded with lint and dust from two production cotton carding machines. The air drawn through the filter and passed through the test tunnel contained fine dust that was similar to the prefiltered air from the model card.

Samples of dust from each of the two dust sources were collected for testing. They were dispersed in an electrolyte and analyzed for size on a volume basis. Model card dust had a mass mean diameter of 3.6μ with a geometric standard deviation of 1.6u. The cotton dust released by tapping the loaded filter varied from 4.5 to 6.7μ mass mean diameter with a geometric standard deviation of about 2. All test concentrations were in the range of 0.5 to 1.5 mg/m^3, which is typical of cotton mill fine dust concentration.

Table VI contains the operational data and collection efficiency for a typical configuration tested.

TABLE VI

EFFICIENCY OF TYPICAL FLUID ELECTRODE PRECIPITATOR CONFIGURATION

Velocity	Field Strength	Dust Supply Concentration	Count Efficiency 1.5u-10u	Mass Efficiency
ft/min	kV/cm	mg/m^3	%	%
250	6.2	0.8	75	86
350	5.9	0.5	81	93
500	4.9	0.8	60	78

In the design, development, and testing of model fluid electrode devices, some qualitative findings are notable concerning fine cotton dust collection or other dust collection. The establishment and control of fluid electrodes requires careful balancing of flow rates and use of smooth tubes to assure laminar flow in the fluid electrodes. In practice, fluid can be piped from a central reservoir through a branching network to fluid electrode locations. Units of practical size would require intermediate flow tubes at approximately 1 ft intervals of vertical fall to catch fluid, decelerate it, and return laminar conditions.

Once properly adjusted, an array of fluid electrodes can operate to continuously remove dust, with low maintenance. Dust entrained in the fluid can continue to circulate through the system until its accumulation warrants a fluid cleaning cycle. Fluid cleaning would not require precipitator shutdown. Applied to textile mills, continuous fine cotton dust removal of 85% would greatly reduce levels of fine airborne dust. For other industrial dusts, possibly with collection fluids other than water, similar applications could be developed.

Further research is being conducted to apply findings to large prototype units, to increase efficiencies with deeper or more optimally selected electrode arrays, and to extend principles to other industrial situations requiring continuous fine dust collection.

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1/ Presented at the Dust Workshop, Notre Dame University, Notre Dame, Ind., April 20-21, 1977.

2/ One of the facilities of the Southern Region, Agricultural Research Service, U.S. Department of Agriculture.

3/ Employee of Battelle Columbus, Columbus, Ohio.

4/ Names of companies or commercial products are given solely for the purpose of providing specific information; their mention does not imply recommendation or endorsement by the U. S. Department of Agriculture over others not mentioned.

WRITTEN DISCUSSION

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Buffalo, New York, New York 14225

The authors should be congratulated for the innovative designs which are described in their paper. These new devices have been designed for the removal of respirable dusts in the textile industry, but could equally well be applied to scores of other applications.

A general comment concerns the reported efficiencies of the three novel devices. The authors state that the indoor dust content of textile mills can be reduced to one mg./m^3, but do not indicate the dust collector efficiency required to achieve one mg./m^3. Therefore, one does not know if the devices described are satisfactory for this service. The most efficient conventional device, the baghouse, is used extensively in the asbestos industry and typically achieves 99.99% efficiency on fibrous dusts less than 15 microns. I do not know why the devices achieving 95%-99% efficiency as described in the paper are considered improvements.

A few specific comments, regarding data and hardware can be made at this point. Two devices utilize corona wires, and I would think these wires would require a method for cleaning deposits which are almost sure to accumulate over periods of time. The Wet-Wall Electronertial unit is described as having low operating costs, but the corona input is approximately 500 watts/1000 CFM. Including perhaps a 1-inch pressure drop and water pumping, it seems that power consumption is relatively high.

According to Table III, in some cases the efficiency of the High Speed Rotary Drum actually decreases when the 1-inch diameter holes are covered with fabric. According to Table VI, the efficiency of the Fluid Electrode Prescipitator increases when velocity is increased from 250 to 350 FPM and field strength is decreased from 6.2 to 5.9 KV/cm. The data in Tables III and VI is contrary to general principles and should be examined more closely.

AUTHORS' WRITTEN RESPONSE

The level of "one mg/m^3" quoted from the paper refers to total dust of all sizes. The efficiencies of present cleaning systems used in textile mills to produce these levels can be greater than 90% for total mass removal of lint, trash, and dust. However, these systems have very poor efficiencies (less than 50%) for capturing respirable dust (less than 15 μm).

Regarding the question of possibly using the baghouse, it is our understanding that the textile industry has had some experience here and that the peculiar nature of cotton lint, trash, and dust is such that baghouses are not satisfactory. These deficiencies are particularly in the areas of automatically cleaning and maintaining the bags and in the tremendous areas required by the high air capacity of textile mills. Our experience with cotton dust is that it is significantly different in character from asbestos dust and no valid analogies between the two can be drawn.

The question about cleaning of trash collected on the corona wires is well taken. Our experience with running the two devices in question is that this has not been a problem. From the standpoint of the W.W.E. unit the vibration which the wire experiences in operation tends to keep it clean. As regards the power requirements, our understanding is that a conventional precipitator could require 50 to 500 watts/1000 CFM with the air velocity below 500 ft/min. The W.W.E. unit operates at 500 watts/1000 CFM but handles air velocities between 2000 and 3000 ft/min. Another advantage of the W.W.E. unit is that its collection surface is about 10 ft^2/1000 CFM, whereas conventional precipitators require from 50 to 1000 ft^2/1000 CFM. These factors (higher operating velocities and smaller collection areas) put the W.W.E. in a very favorable position compared to a conventional precipitator or other air cleaner.

With regards to the Reviewer's comments concerning anomolies in the efficiency of the rotary drum as given in Table III, it is true that in some instances the efficiency of the drum did decrease when the 1-inch diameter holes were covered with fabric. This has been explained by the fact that apparently the particle counter was registering particles that were being knocked off of the back side of the filter fabric. Concerning the apparent problems posed in interpreting Table VI, it should be clarified that this data is not presented for all the same experimental configurations, thus explaining the reason for the seeming inconsistency in the way the efficiency followed the principles of physics. This data was merely presented as a sample of the types of efficiencies obtained with the fluid electrode precipitator.

OPEN DISCUSSION

Baril: As far as we know we do not have to clean the corona wire. We have not had any problem on the wet wall electro-inertial unit. The rotating drum was looked at as an air cleaner , also as an electro-inertial unit.

Reif: For the rotary drum the dust goes in the bottom of the box, most of it does not enter the holes in the drum. The comment about the fiber filter not seeming to improve the effectiveness of the device is true. These types of fiber filters are used currently in the textile industry which do not have very high efficiencies. That is one reason why we were looking at them. If you look at them with the atmospheric dust type test that we were using, their efficiencies are just a few percent by themselves. The base velocities are about 200 ft/min. So they are very poor filters actually. The first stage filters were designed to pull out long cotton fibers, the second stage filters are generally of higher quality but still are not particularly good filters.

Baril: Concerning ourselves with table 2, the artificial cotton dust at zero potential and zero current has an efficiency percent of about 80.2%. Then right under it is a card output (this is at zero current), its efficiency is 30.3%. The content of larger particles is different and you are seeing the effect of electro-inertial input.

Billings: I do know you had card type dusts and this analysis is a serious problem. I am concerned what you will do with this collector and how you will distribute the dust, and what you are talking about in terms of the actual staple, yarn and textile weaving. Are you basically going to have a ventilation system which sucks all of the air out of one end of the building, cleans it and redistributes it? Are you going to to have local exhaust ventilation on every dust source? How are these going to be configured in attempting to meet the standard?

Baril: You have asked a question that people in the air handling textile industry suggest that the best thing they could do is get out of that industry and service some other industry.

This is a very serious problem and the people who service the textile industry are few and far between (one is Albert Baril). To answer this question let us go back to a question asked by Dennis Helfritch. He said you have units and this and that, we readily agree that you have a lot of units and they work very well. We could go to a bag unit which is very efficient. EPA has looked at some of this. The estimate comes back that the baghouse must be about the size of the textile mill. This is one of the problems.

Right at the moment, we do not know what we are going to do. The government is going to set some standards. The Department of Agriculture is going to try like mad to get the standard high. If we get that, we are not in as much trouble as you propose. If we do go to the 200 bin it will take us from 5 to 7 years to get there. If we set a standard that will bring it down to 200 in 5 years, who is going to buy a piece of machinery that will bring it down to 500 in a year?

What it looks like right now is that a lot of the weave rooms have air which comes in at a high level and escapes completely through the floor and is taken out and cleaned or it sweeps across the room and does this. To increase that system to handle what the bin could possibly take, I then think we might have to have terrific exchanges of air. We would have to bring this air down to someplace where we could actually clean it. This incurs a cost exceeding $2 billion. This was worked out by TRI Princeton. Right now this section is really up in arms. One possible way to go is to build individual containment units around many machines. They have already been successful in doing this on the other end of the textile industry; that is cards, enclosing them and capturing them.

We also enter a second problem which we have not addressed at all and that is capturing the dust so you can clean it. You do have leakage into the system and you can only capture a percentage of the total dust, and by cleaning, putting it back into the room, you are also leaking some of the unwanted particles into it. The capture system has to be worked on too.

We have been working on an entirely, completely new system for textile mills which puts all the systems together in an integrated machine. In other words, you start off at one end by putting in cotton puffs and you get yarn out of the other end of the machine. This machine could be totally enclosed. We are doing research on that line.

Billings: It is important to mention one other federal agency which comes into the discussion and that is the FDA. If you use an electro-static precipitator that involves any kind of ionization you are going to make ozone. There are already ozone standards with regard to this kind of technology.

Baril: We have been monitoring the ozone levels.

Benarie: I have some information about rotary filters that are used, for instance, under a British patent on submarines. There are three possibilities for building rotary filters. The mechanism of the rotary filter is quite easy to explain. The impaction mechanism on the filter, on the individual fibers, instead of being impacted individually by the vector of the air stream is impacted by the vector of the whole movement within the airstream which can be 10 times as large, at say 3000 rpm. The fibers can be oriented either radially or longitudinally and the stream can be from inside toward the outside or in reverse. Now the reverse system seems much better because the cake is thrown away automatically by centrifugal force. Rotary air filters are made which are used in the electro-chemical industry and are quite good.

I have some more information about it if you are interested.

Baril: We built a model that the British have and it was designed primarily to remove radioactive particles from the stream. We built 3 units and approached it 3 ways. All three were good, but we felt the fluid electrode was the best. We have not given up on that. We still have it and work on it.

DIELECTROPHORETIC AIR FILTRATION: PROGRESS AND PROBLEMS

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ABSTRACT

Dielectrophoretic air filters under development at the Naval Research Laboratory, as well as those previously reported or currently manufactured, are discussed with a view to establishing successful design principles.

INTRODUCTION

Our burgeoning air and gas filtration technology concentrates largely on the removal of particulate material from the gaseous discharges of industrial processes. This is legally necessary and highly desirable from an environmental health point of view. There is, however, another air filtration challenge: that of cleaning the air drawn into, and circulated within, occupied buildings, vehicles and other enclosures. The aerosols of principal concern are of several types: (1) the airborne pathogens that can cause cross-infection in homes, hospitals, schools, theaters, etc.; (2) the miscellaneous airborne debris that affects the operation of delicate electrical and electronic equipment of many kinds; (3) the common dirt and dust that deposits on surfaces and requires control as a matter of esthetics and good housekeeping; (4) the particulates formed in motor vehicle operation; and (5) the industrial airborne refuse which is not removed at its sources. Other common aerosols may in time be thought sufficiently objectionable to require removal from the ventilation air of buildings. Such aerosols include the sulfate/sulfuric acid originating in the burning of sulfur-containing fuels (1), the terpene-derived blue haze produced by vegetation (2), and the metal-containing particles produced by vegetation (3).

The means presently available for high-efficiency filtration of ventilation air appear to have sufficient disadvantages to severely limit their use. Such means include HEPA filters, very high quality glass fiber media, and electrostatic precipitators. Their disadvantages are either high cost or high pressure losses, or both.

We believe that a promising means of handling present and future problems in ventilation air filtration is the technique of dielectrophoresis.

Equipment of this type is inherently simple, low in cost, and very compact. Since a dielectrophoretic air filter is a field-operated rather than a current-operated device, its power consumption is near zero. Additionally, there is no corona discharge, ozone production is zero, and sparking between electrodes can be eliminated.

A brief historical review of dielectrophoresis and dielectrophoretic air filtration is desirable, both to give credit where due and to assist others in avoiding duplication of ineffective or non-optimized designs.

The word dielectrophoresis was coined by Pohl to describe the accelerated separation of solids from liquids in nonuniform electric fields (4). We have continued to use the same word to describe the accelerated separation (filtration) of solid or liquid particles from air in nonuniform electric fields. Such an unusual descriptor helps nonspecialists to understand that we are not talking about electrostatic precipitation.

Let us first describe and discuss briefly the dielectrophoretic process. Dielectrophoresis is the movement of a particle in a nonuniform electric field, the direction of movement usually being toward the most intense part of the field. This mechanism is conveniently applied in air filtration by passing the air through a thin layer of air filter material which is sandwiched between metal screen electrodes. The air moves through the sandwich perpendicular to its plane, and experiences no resistance other than that of the filter medium itself, which may be a commercial glass fiber mat or a slab of open-pore plastic foam. Since every glass fiber or every element of the foam locally distorts the applied electric field, the mat or slab contains an enormous number of field inhomogeneities. Since the glass fiber and the plastic foam have dielectric constants greater than that of air, the field is more intense close to a fiber or foam element than remote from it. Therefore airborne particles within the filter medium experience a dielectrophoretic force. This force is approximately described by:

$$F = 2 r^3 K_1 \left(\frac{K_2 - K_1}{K_2 + 2K_1} \right) E^2$$

Thus, the dielectrophoretic force F is proportional to the square of the applied field E and, to the cube of the particle radius r; the dielectric constants of the medium, K_1, and of the particles, K_2, are also involved.

NRL STUDIES

The NRL studies have employed a uniform geometry comprising a flat glass fiber mat, 0.64 cm thick, separating two steel wire screen electrodes. Variables have been the composition of the mat (three as-received commercial air filter media), voltage (0 to 7kv, equivalent to 0 to 11 kv/cm), air velocity, and aerosol (DOP of 0.3 and 0.8 micrometer diameter, and a fine fly ash). The dielectrophoretic effect is illustrated by the Dielectrophoretic Augmentation Factor (DAF). This is the ratio of the aerosol percent penetration (100% - percent aerosol retention) at zero voltage to the percent penetration at the voltage of interest. Tables 1 and 2 present data for one filter medium (the Farr Co's HP-100) at several voltages and air speeds, and for two DOP particle sizes. The manufacturer's rated air speed for the HP-100 medium is 20 cm/sec (40 fpm), at which speed the initial resistance is 1 cm wg (0.40 iwg). The DAF's at 7 kv and 20 cm/sec are 21 for 0.3 micrometer DOP and 28 for 0.8 micrometer DOP.

TABLE 1

DAF VS AIR SPEED AND VOLTAGE FOR HP-100 FILTER MEDIUM;
AEROSOL 0.3 MICROMETER DOP

Air Speed cm/sec	Applied Voltage, Kv 2	3.5	5	7
3	8	19	95	330
5	3	13	39	120
8	3	11	28	100
13	2	6	13	42
18	2	5	9	27
26	2	4	6	14
37	2	3	4	9
46	1	2	3	6

TABLE 2

DAF VS AIR SPEED AND VOLTAGE FOR HP-100 FILTER MEDIUM;
AEROSOL 0.8 MICROMETER DOP

Air Speed cm/sec	Applied Voltage, Kv 2	3.5	5	7
3	30	110	300	1100
5	6	30	95	360
8	4	18	50	170
13	3	10	20	50
18	2	6	13	35
26	2	4	8	18
37	2	3	5	11
46	1	2	3	7

It should be noted that the dielectrophoretic force does not operate alone in a filter medium. There are also the ever-present aerodynamic forces and, in some cases, forces due to net electric charges on the particles, the filter medium, or both. Thus, a complete analysis of the performance of dielectrophoretic air filters has not yet been made. It is clear that the relationship between the dielectrophoretic force and the applied field squared is likely to hold regardless of aerodynamic factors. However, since both the aerodynamic forces and the dielectrophoretic forces are affected by both particle size and filter element size, the total interaction may be difficult to predict.

We offer the hypothesis that the dielectrophoretic force is effective only at shorter range than the aerodynamic forces and that therefore, the former is operative only when there is pronounced evidence of aerodynamic effects in the absence of the electric field.

The interplay between the aerodynamic and the dielectrophoretic forces is complex. For example, aerosol particles of a given size which are moving through a filter medium deposit inertially on a small filter element more readily than on a large one. However, the electric field gradient is stronger adjacent to a large filter element than to a small one. There would seem to be a conflicting requirement here: small filter elements for inertial deposition of particles, and large elements for dielectrophoretic deposition. In fact, however, small elements seem to be desirable for dielectrophoretic deposition also; only with small elements do the aerodynamic effects allow particles to pass close enough to the elements for the dielectrophoretic foam to act effectively.

OTHER DIELECTROPHORETIC AIR FILTERS

As far as we know, the first dielectrophoretic air filter, the "Electro-Polar", was constructed by Sproull and others at the Western Precipitation Corporation in Los Angeles during the early 1950's (5). This was a full-scale device with an air flow rate of 3500 cfm. The two filter media used were typical commercial glass filter materials of good quality. It was tested by the Harvard University Air Cleaning Laboratory for the Atomic Energy Commission. The reported results were mediocre (6,7), although in our view the performance was better than reported. The dc voltage used for the Electro-Polar device was 15 kv for a one-inch medium. This provided a gradient of 5.9 kv/cm. Influential in the choice of voltages was the desire to avoid both corona discharge on the expanded-metal electrodes and a spark discharge through the filter medium.

A report by Thomas and Woodfin in 1959 described a dielectrophoretic air filter of unusual design (8). The spaces between a series of charged and grounded metal plates, oriented parallel to the air flow as in an electrostatic precipitator, were packed with glass fiber. A filter one inch deep in the direction of the air flow showed "great improvement" when the voltage was applied. The performance of a filter 14.5 inches deep at a voltage gradient of 9.3 kv/cm was improved by a factor of over five at air velocities of 400-500 fpm. No further mention of the Thomas-Woodfin device is found in the literature.

Havlicek's valuable theoretical dielectrophoresis paper of 1961 reported on the effectiveness of electric fields parallel and perpendicular to the direction of air flow (9). The former was found to be preferable on the basis that it allows the maximum field gradient to occur at points where the relative particle-to-fiber velocity is at a minimum. Thus the two effects act together to maximize particle deposition. In fields perpendicular to air flow, deposition is correspondingly minimized. Havlicek's theorem indicates that the geometry of Thomas and Woodfin's filter was not such as to optimize filter efficiency.

Rivers' 1962 paper provides a useful theory review and, in addition, describes a unique air filter which may have some dielectrophoretic properties (10). The medium is a multilayer paper; the electric field causes a separation of the layers. The modest improvement due to electrification has been attributed to separation of the paper layers, probably attributable to electrostatic repulsion. Rivers' device also places the electric field perpendicular to the air flow direction.

The next example in our historical review is that of Walkenhorst and Zebel in 1964 (11). Their filter medium unit was ten layers of nylon hosiery fabric between screen electrodes; several of the units were placed in series. Using aerosols of crushed coal or quartz, good filtration performance was obtained in an electric field. Remarkably, a similar filter, exposed to the NRL DOP aerosols, showed no significant particle deposition.

Finally, a currently manufactured air filter of dielectrophoretic design is the "Fiberstatic", by Filterlab of Houston, Texas. The design of this device is a high voltage screen electrode having a 0.6 cm polyurethane foam mat on each side, all sandwiched between two grounded screen electrodes. In a test with a DOP aerosol no appreciable filtration efficiency was observed.

My colleagues and I would now like to discuss the historical cases we have cited in order to assist future development of dielectrophoretic air filters. We consider that the Electro-Polar of Sproull, et al was the best of those we have reviewed, as well as being the earliest. Why, then, did not the Electro-Polar perform in consonance with the potential indicated by the NRL data. Perhaps it was an idea whose time had not yet come. More significantly, the problem may have been the unavailability of glass filter mats thinner than 1.3 cm (0.5 in). The thicker media available required a higher voltage to attain an efficient voltage gradient, but the higher voltage increased the probability of the undesirable corona discharge. Forced to use a low voltage gradient, the Electro-Polar team then were hurt by the square-law dependence of dielectrophoretic force on voltage gradient. Another factor may have affected the operating voltage chosen. This was the increased probability of a spark discharge through the filter medium as the voltage was raised. Spark discharges produce holes in the filter mat, and may ignite flammable materials deposited in the filter. It is possible that the sharp edges of the expanded metal lath electrodes were likely sources of both corona and spark discharges. This problem has not yet been fully solved, but an NRL development may eliminate this difficulty. In any case, steps should be taken in dielectrophoretic filter design to allow operation at the highest practicable voltage gradient, taking advantage of the square-law relationship.

Turning attention to the Rivers and Thomas/Woodfin devices, we believe that they demonstrate the validity of Havlicek's axiom that dielectrophoretic filters should be designed for parallel alignment of air flow and electric field. This should be considered an established principle of dielectrophoretic filter construction

Walkenhorst and Zebel's filter, with the stacked nylon hosiery medium, has the property of good performance with solid aerosols of crushed coal or quartz, and negligible performance with the NRL DOP aerosols. Further, the relatively coarse nylon filaments (10 to 15 micrometers) would not be expected to be well suited as dielectrophoretic filter fibers. We suggest reconciling the Walkenhorst/Zebel and NRL data as follows. The nylon medium per se is in fact ill suited for dielectrophoresis because of its large fiber diameter. However, as a few solid aerosol particles deposit on the fiber, probably by an aerodynamic mechanism, the sharp points and edges of the deposited solid particles serve to produce electric field inhomogeneities. Dielectrophoretic action then increases progressively. With DOP aerosols, some DOP droplets may also deposit aerodynamically, but then yield no sharp points or edges. In fact, they simply spread smoothly over the nylon fibers, increasing their diameter slightly. If the above reasoning is correct, widely differing performance can be expected from a large-fiber or large-element dielectrophoretic filter, depending on whether the test aerosol is a liquid or certain solids. It is possible, also, that the satisfactory performance of Walkenhorst/Zebel filter with the coal or quartz aerosols was in part dependent on particles larger than the 0.3 and 0.8 micrometer DOP.

While we have no test data on the Fiberstatic filter with a solid aerosol, it is plausible that this filter performs well in the manufacturer's hands, and that the test aerosol is a solid. Thus, the test pattern would correspond to that of the Walkenhorst/Zebel filter. This could be expected since the elements of the reticulated polyurethane foam are also very large compared with the glass filaments of the NRL filters.

CONCLUSIONS

We believe that dielectrophoretic air filtration has valuable characteristics that will contribute toward very wide application in the future. However, even the "first generation" NRL filters require careful consideration of the basic phenomena to insure satisfactory performance. Consideration of the design of prior and current dielectrophoretic air filters is helpful in producing new designs.

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WRITTEN DISCUSSION

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In accepting this assignment reluctantly, I reminded Professor Ariman that my experience has been confined mostly to the practical aspects of particulate control by fabric filters without electrical augmentation. While quite sometime ago, limited studies were conducted with artificial electrification of an experimental fabric filter[1], most of my work has been restricted to considerations of the effects of natural charging in both the media and particulates. But because of the very outstanding benefit realized by augmenting the charge on the medium and my personal views concerning critical importance of electrostatic involvement in the filtration process, I can be completely sympathetic to the Thompson et al observations. Hopefully, studies such as this will receive further attention.

I would like to take this opportunity to ask the following questions that, perhaps, I should have had answered from my own earlier evaluations if only they had been made under more controlled conditions at that time.

1. How was the aerosol generated and conducted to the filter and was it charged in the process?

2. What, for example, is the ultimate fate of the collected product?

3. What form does it take as it deposits or forms on the collecting surface(s)?

4. How much longer can the electrified field filter operate than the non-electrified variety?

5. Since we might expect DOP to coalesce, particularly in the field, does this occur and to what extent?

6. If the DOP mist coalesces on the collecting fibers, does the oil form on - and then run from - the surface without affecting, adversely, the filter's performance for extended periods of service?

7. If this is a realistic appraisal of the process, can it be applied to other mists or mist-like emissions? And if so,

8. What are the limitations, considering the flammability and explosiveness of some mist-like solvent emissions?

The above questions are based primarily on my suspicion that, like effective natural charging, electrical augmentation may lead to improved aggregation or agglomeration of the collected product. Most often, when this is accomplished ideally, the nature of the deposited cake is most favorable for optimum collectability, for most effective cleanability and for highest efficiency. With a condensable liquid mist, this effect should be easy to detect.

Professor Penney has provided significant substantiation of some of our experiences through his fundamental studies by relating the combined effects of particle charge and fields with the deposited product. He has shown that "impact" charged particles form a chain-like deposit on a fabric without the use of any high voltage on the particulate or on the collecting surface. In another study, again without external potential impressed on the filter fabric, corona-charged particles also deposited in a chain-like aggregated manner on one and only one of the fibers in a composite, two fiber filters.[2]

Figure 1 is a photomicrograph showing a blend of two different (wool and acrylic) fibers about 3 μm in diameter. The fact that one fiber remains clean while the adjacent fibers and only these electropositive fibers collect the negatively charged particles as a porous aggregate is extremely significant. If this is the kind of open deposit that can develop because of electrostatic forces, it would seem likely that the exceptional performance of the NRL dielectrophoretic filter could also be attributed to the formation of such a super porous deposit.

In the simulated filtration tests that we carry out without charge augmentation, natural electrostatic charges, when suitably balanced, offer a most

FIGURE 1 AGGREGATED DEPOSIT OF PRECHARGED PARTICULATE ON A FILTER FIBER

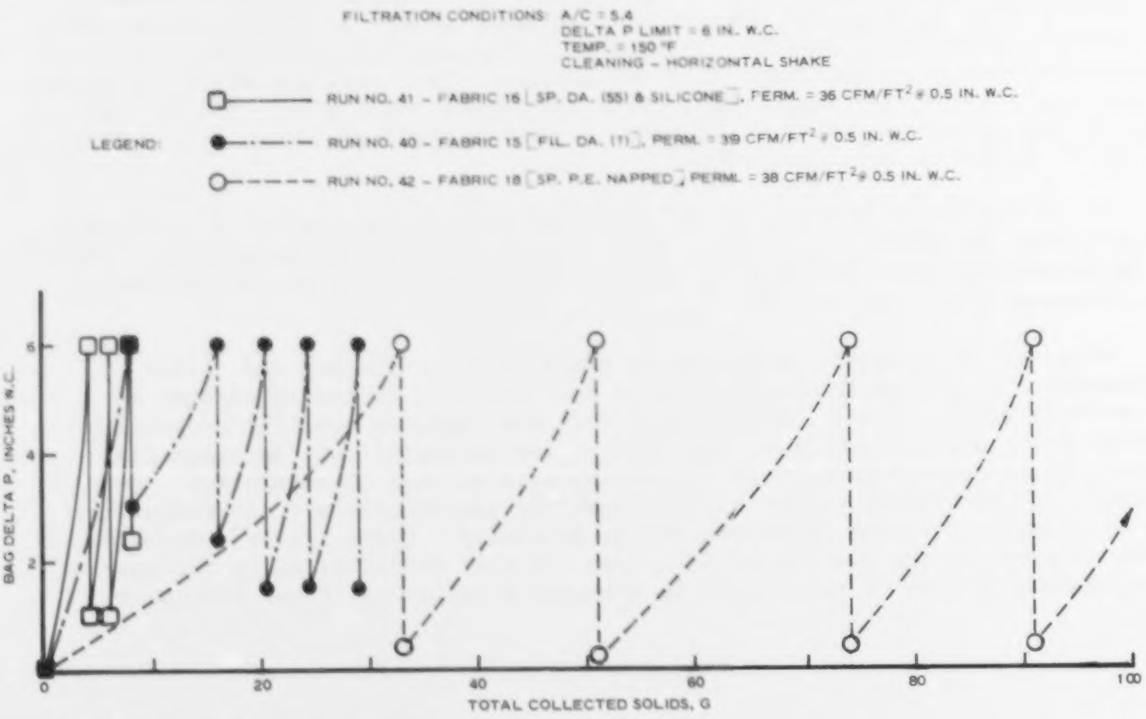

FIGURE 2 EXPERIMENTAL FILTRATION OF ELECTRIC FURNACE DUST

desirable means for achieving the highest level of filtration performance. The evidence implies that when a particulate will form a porous deposit or aggregate, this most desirable change in the cake can be achieved during the filtration process by a favorable balance of the electrostatic properties of the particulate and of the collecting fabric. While most particulates can form porous deposits, not all dusts aggregate so easily that any but the more active (electrostatically) fabrics can promote the change. Extraordinary persuasion seems possible by using media of the most suitable electrostatic polarity, charge intensity and, sometimes, of perferred charge dissipation rate.

On page 12 of the Helfritch paper "Electrical Filtration and the Apitron/ Design and Field Performance," results are given for pilot tests on electric furnace (steel) dust. The curves indicate that electrical augmentation increases filtration performance (rate and efficiency) very significantly. The curves obtained in our non-electrified filtration tests also carried out on electric furnace (steel) dust have a similar appearance indicating, possibly, the same kind of improvement in deposited cake. Thus, Figure 2 shows a plot of the quantity of electric furnace solids collected as a function of bag pressure drop (limited in the tests to 6 inches w.c.) for three polyester fabrics used in the writer's filtration study of an electric furnace (steel) dust. The conditions in the three evaluations were essentially identical and the fabrics, although of different construction, were of about the same permeability. Significant variations were noted in electrostatic properties among the three media with the filament Dacron fabric (#15) being quite electropositive, the napped polyester fabric (#18) being most electronegative with highest intensity and the spun Dacron fabric (#16) of intermediate polarity and intensity.

Professor Penney's reproduction of nodules or "dingle-berries" in experimental filter tests also seem to reaffirm the fact that charges play a very important role in aggregate formation. The nodules were produced on high cover (fuzzy), not filament (smooth), fabrics with a strong electric field. In subsequent tests, it was found that the nodules formed more quickly in a region where there was little field compared to the strong field area. Nodule formation can be induced by electrostatic forces and their retention to the fiber seems to be due to this same force.

It is clear that not all of the facts concerning charge related particulate aggregation are known, but it does seem evident that the porous deposit - low flow phenomenon resistant type of cake formation could account for the super performance of the NRL filter.

Whatever the specific mechanism by which electric charges and fields influence the collection process, it is apparent that if the advantages of electrical effects are to be generally utilized, certain specific properties of filter media and particulates, not now considered significant in commerical fabric filtration circles, need to receive more serious consideration. Certainly, the electrical resistivity of both the particulate and the medium have to be considered worthy of determination routinely. Quite likely, oversight of this property in the past could have been a factor contributing to the poor or non-reproducibility of data noted in Thompson's review of "Other Dielectrophoretic Air Filters."

That high resistivity is a problem limiting the charging of certain fly ashes and, thereby, restricting their collection by electrostatic precipitation, is well established. It is also clear that the resistivity of other particulates vary from relatively low (10^8 Ω cm) to extremely high (> 10^{14} Ω cm) values. Since particulates and fabrics of high resistivity cannot be charged, except naturally, it is quite possible that some inconsistencies in performance and the non-reproducibility of results indicated with the various dielectrophoretic devices reported by Thompson et al can be attributed to such variations.

Our experience with many filter fabrics provided by a number of different suppliers indicates conservatively, that up to 20 percent of the commercial media offered to the baghouse users carry residual finishes which reduce the resistivity of the fabric to 10^{10} Ω/□ or less. It is not too difficult to imagine, then, that different lots of supposedly the same product have performed differently in dielectrophoretic trials, just as they function differently in non-electrified filtration operations.

Product differences among manufacturers introduce another variable and represent a problem that is more the rule than the exception. When to this uncertainty is added the strong competition to cut prices and, in some instances, a policy of interchanging fibers of a generic type under one specification, it becomes evident that reproducibility in the performance of filter media cannot be achieved without identical duplication.

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OPEN DISCUSSION

Penney: There was quite a flurry of activity among manufacturers on this about 25 years ago. They were trying two-stage precipitators. Nothing was ever published about it. There was a big difference whether you could charge the corona dust, then you could get fairly high efficiency. But you were in trouble if the efficiency was not high enough because you would have space charge troubles.

There was the remark about a filter giving a high efficiency on quartz or coal dust and not on DOP. That is to be expected. My experience has told me that it is difficult to disperse the quartz without having the particles strongly charged.

The same thing happened with Rivers when he presented the paper at Penn

State. He noticed that his artifically dispersed dust gave a high efficiency, but the atmospheric air which has had a long time to get naturally discharged showed a low efficiency. It seemed a question of whether or not your test particles are charged or not.

Frederick: Do you think that the fact the particles were solid rather than oil spheres contributed to their collection efficiency?

Penney: I do not know. I think it is more than likely to be the dispersion of particles.

Friedlander: This paper is quite interesting; it shows the similarities and differences from the HGMS. First of all Zebel's paper is not quoted by the authors. This paper shows very clearly that it is easy in the electric case to get confused between the dipolar and the charged cases. You can not keep them apart. In the magnetic case, there is no confusion because there are no monopolar magnetic charges available.

In this case you have the problem of electric dipolar. Magnetics people do not have magnetic monopoles because we do not have monopoles. We need large currents of low power if we use superconducting currents. Many applications would be well suited for permanent magnets. The ferro-electric does not really get high enough electric radiation.

I think that the electro-ferretic people can learn from HGMS. Because all the things discussed in this paper have been thoroughly investigated in HGMS. Now you could go back and see where you could use this information the best.

COLLECTION OF FINE PARTICULATES USING A FOAM SCRUBBER

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ABSTRACT

This paper addresses the subject of application of foam scrubbing to fine particulate control. Diffusion and sedimentation appear to be the primary mechanisms of particle collection. Fundamental theory for these two mechanisms is discussed.

Principles of foam scrubbing were experimentally verified for particles between 0.18 and 1.0 μm diameter on a bench (2.5 acfm) and small pilot scale (500 acfm). The results of these experiments are presented. In either scale the experimental results agree well with the theory. Comparisons of economics indicate that the foam scrubber can be competitive with conventional collection devices including electrostatic precipitators, fabric filters, and high energy scrubbers in both capital investment and operating costs.

ACKNOWLEDGEMENT

This work was supported by the EPA under contract 68-02-1453. Appreciation is expressed for EPA's continued interest in the foam scrubber.

INTRODUCTION

The basic principle of the foam scrubber is illustrated in Figures 1 and 2. Gas stream contaminated with the particulate matter is forced through a screen wetted with a surfactant solution. As the gas passes through the wet screen openings, foam is generated and the gas is enclosed in a multitude of small bubbles, Figure 2. The distance which a particle must travel to reach a collection surface is now greatly reduced by the bubble enclosure. The foam

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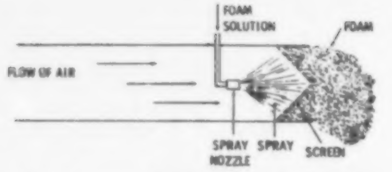


Figure 1. The basic principle of foam scrubbing

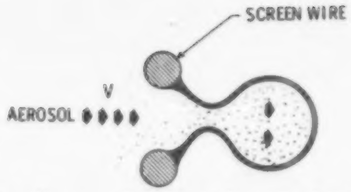


Figure 2. Aerosol enclosure in a foam bubble

proceeds through the main chamber of the scrubber where the particles are given time to migrate to collide with the bubble wall. As the foam is broken, particles retained on the bubble walls are collected in the surfactant liquor.

Monsanto Research Corporation (MRC) initiated work on foam scrubbing in the latter half of 1974 under a contract with IERL-RTP, EPA. The contract involved theoretical evaluation, a bench-scale verification, and a small pilot scale demonstration of the foam scrubber. Results of this work form the basis of this paper.

The next section deals with theoretical aspects of particle collection by foam. The results obtained on a bench-scale (2.5 cfm) and small pilot scale (500 acfm) foam scrubber units are then presented. Finally, the economics of the foam scrubber are presented.

In this study, a scrubber system using wet screen foam generation was considered (see Figure 1). It should be noted that other methods of foam generation may result in conclusions different from those stated in this paper. Methods of foam generation other than the wet screen are not a subject of this paper.

THEORETICAL ASPECTS OF PARTICLE COLLECTION BY FOAM

Several governing equations could be applied to describe the process of foam scrubbing. These include the equation of continuity, momentum (Navier-Stokes), energy, and diffusion. As is often the case in practical situations, exact descriptions of the phenomena using these equations are characterized by rigorous formulas too difficult to solve mathematically. As a result, many simplifications must be applied.

Let us first consider the idealized case of bubble formation as shown in Figure 2. Aerosol with constant particle concentration, C_0, rushes into a foam bubble as it is being formed. The bubble containing the aerosol detaches at time $t = 0$. As long as the cross-sectional areas of the foam generator and the foam scrubber are similar, the mass of aerosol stream will move with essentially the same velocity before and after the formation of a bubble. It is therefore reasonable to assume that the aerosol inside the bubble and the foam bubble surface progress with the same velocity through the scrubber. Consequently, the impaction forces active in this system may be assumed to be negligible.

Now let us consider diffusion of particles enclosed in the foam bubble. Assuming an isothermal system (i.e., gas temperature - bubble film temperature) and a very small, spherical bubble, if the motion of gas is negligible inside the bubble after the bubble has been formed, then from the set of governing equations, Fick's Second law applies:

$$\frac{\partial C}{\partial t} = D\nabla^2 C \qquad (1)$$

Assuming uniform aerosol concentration and transformation of Equation 1 into spherical coordinates, we obtain:

$$\frac{\partial C(r,t)}{\partial t} = D\left[\frac{1}{r^2}\frac{\partial}{\partial r}\left(r^2\frac{\partial C}{\partial r}\right)\right] \qquad (2)$$

The diffusion coefficient, D, for small particles may be expressed by[1]:

$$D = \frac{k'T}{6\pi\mu r_p}\left[1 + A\left(\frac{\lambda}{r_p}\right) + Q'\left(\frac{\lambda}{r_p}\right) e^{-br_p/\lambda}\right] \qquad (3)$$

where k' = Boltzmann's constant (1.38×10^{-22} kg m/s^2 °K)

T = temperature

μ = viscosity of gas

r_p = radius of diffusing particle

λ = mean free path of the gas

A, Q', b = empirical constants

The solution of Equation 2 will give the particle concentration distribution inside a bubble as a function of time and radius for various particle sizes, assuming that the particulate cloud is monodisperse and, at given concentration, the coagulation of particles is negligible.

The partial differential equation, Equation 2, can be solved by the separation-of-variables method. To solve Equation 2 the initial and boundary conditions must be summarized:

(a) $C(r,0) = C_o$ initial concentration at $t = 0$ is constant (4)

(b) $C(R,t) = 0$ on the surface of the sphere at $t > 0$ (5)

(c) $C(r,t) \neq \infty,\ r \longrightarrow 0$ (6)

The general solution will have a form as follows:

$$C(r,t) = (C_1 e^{-\alpha^2 Dt} + C_2 e^{\alpha^2 Dt}) \left[C_3 \frac{1}{\sqrt{r}} J_{\frac{1}{2}}(\alpha r) + C_4 \frac{1}{\sqrt{r}} J_{-\frac{1}{2}}(\alpha r) \right] \tag{7}$$

where $\alpha, C_1, C_2, C_3, C_4$ = constants

J_n = Bessel functions

By use of boundary conditions, Equations 4, 5, and 6, and lengthy derivation and transformation, Equation 7 becomes:

$$\frac{C(r,t)}{C_o} = \frac{2}{\pi} \left(\frac{R}{r} \right) \sum_{n=1}^{\infty} \frac{(-1)^{n+1}}{n} e^{-(\frac{n\pi}{R})^2 Dt} \sin(n\pi \frac{r}{R}) \tag{8}$$

This equation gives the instantaneous particle concentration inside the bubble.

The instantaneous rate of loss of particles to the bubble surface is now given by:

$$q(R,t) = -D(4\pi R^2) \left. \frac{\partial C}{\partial r} \right|_{r=R} = 8\pi RDC_o \sum_{n=1}^{\infty} e^{-(\frac{n\pi}{R})^2 Dt} \tag{9}$$

The total loss of particles in time t from time t = 0 is:

$$Q(t) = \int_0^t q(R,t)dt = \frac{8R^3 C_o}{\pi} \sum_{n=1}^{\infty} \frac{1}{n^2} \left[1 - e^{-(\frac{n\pi}{R})^2 Dt} \right] \tag{10}$$

Hence, the particle collection efficiency, E (%), in time t may be determined as follows:

$$E = \frac{Q}{4/3\pi R^3 C_o} 100 = \frac{6}{\pi^2} \sum_{n=1}^{\infty} \frac{1}{n^2} \left[1 - e^{-\left(\frac{n\pi}{R}\right)^2 Dt} \right] \times 100 \qquad (11)$$

note: $\sum_{n=1}^{\infty} \frac{1}{n^2} \equiv \frac{\pi^2}{6}$

Inspection of Equation 11 shows that bubble diameter, R, is of primary importance to collection efficiency. It appears as a square term in the denominator of the exponential term. For illustration, particle collection efficiency, Equation 11 was plotted for bubble size of 1.0 mm diameter and three particle sizes, 0.01, 0.1, and 1 μm diameter, as a function of foam residence time, Figure 3. Collection efficiency by diffusion only of 0.01 μm particles within a reasonable practical residence time (say 5 seconds) is very high as compared with a 1.0 μm particle.

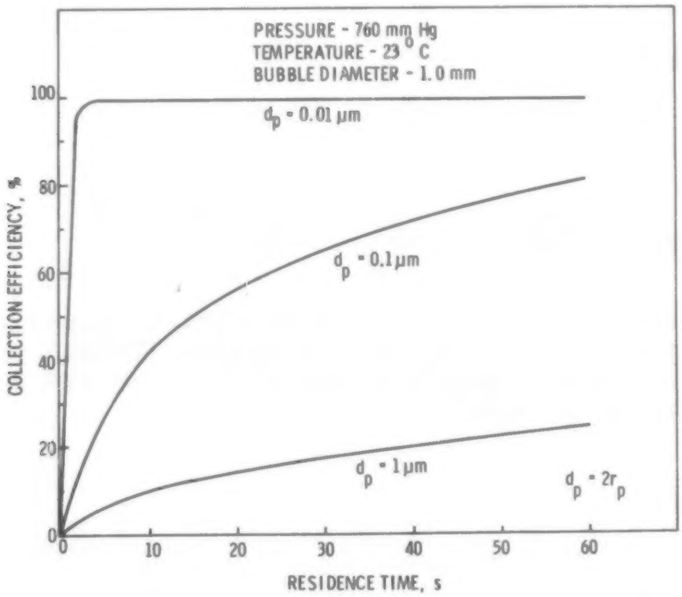


Figure 3. Theoretical particle collection efficiency

It is obvious that some of the enclosed particles could reach the bubble surface by sedimentation. The number of particles deposited per unit time inside the bubbles by sedimentation can be expressed as $(\pi R^2)CV_s$, where V_s = particle settling velocity. For simplicity, it is assumed that the concentration of particles, C, inside the bubble is uniform. Using this expression we can formulate the coefficient of absorption by sedimentation (ratio of the rate of particles deposited to the total number of particles in the bubble), α_s, as follows:

$$\alpha_s = \frac{\pi R^2 C V_s}{(4/3\pi R^3)C} = \frac{3V_s}{4R} \qquad (12)$$

The settling velocity can be expressed by[1]:

$$V_s = \frac{d_p^2 \rho_p g}{18\mu} \left(1 + A\frac{2\lambda}{d_p} + Q'\frac{2\lambda}{d_p} e^{-bd_p/2\lambda} \right) \qquad (13)$$

where ρ_p = particle density

Combination of Equations 12 and 13 gives:

$$\alpha_s = \frac{d_p \rho_p g}{24\mu R} \left(1 + A\frac{2\lambda}{d_p} + Q'\frac{2\lambda}{d_p} e^{-bd_p/2\lambda} \right) \qquad (14)$$

Typical sedimentation velocities of particles (Equation 13) and coefficients of absorption by sedimentation (Equation 14) are listed in Table 1 for a bubble diameter of 1 mm and a particle specific gravity of 1 g/cm³.

Table 1. TYPICAL SEDIMENTATION VELOCITY AND COEFFICIENT OF SEDIMENTATION (for $R = 0.0005$ m, $\rho_p = 1 g/cm^3$)

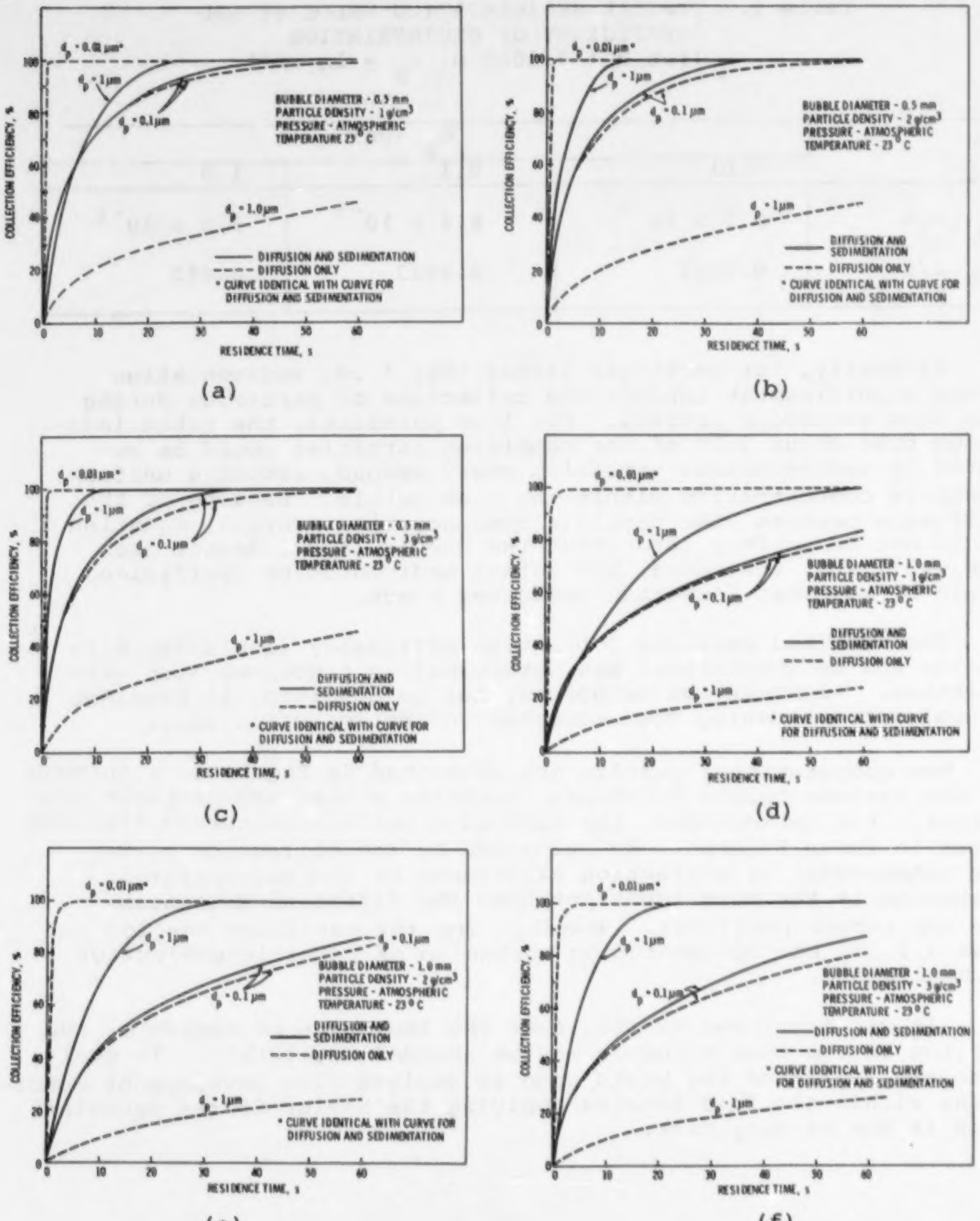
	d_p (μm)		
	0.01	0.1	1.0
V_s, m/s	6.5×10^{-8}	8.8×10^{-7}	3.5×10^{-5}
α_s, 1/s	0.0001	0.0013	0.052

Evidently, for particles larger that 1 μm, sedimentation would significantly enhance the collection of particles during the foam scrubbing process. For 1 μm particles, the table indicates that about 5.2% of the remaining particles could be removed by sedimentation mechanism every second, assuming uniform particle concentration within the foam bubble. Because of the diffusion process, the particle concentration profile, Equation 7 would not be uniform throughout the bubble (i.e., leaner near the wall) and, therefore, the actual sedimentation coefficient would be somewhat less than indicated above.

The combined particle collection efficiency (including diffusion and sedimentation) was determined in a step-by-step calculation. Although the method may not be absolute, it provides a tool for estimating foam scrubber collection efficiency.

The computational results are presented in Figure 4, a through i, for various bubble diameters, particle sizes, and particle densities. For convenience, the diffusion collection curves are also shown in these figures. As evidenced by the collection curves, the enhancement of collection efficiency by the sedimentation mechanism is far more important than the diffusion mechanism for the larger particles. However, for the particles smaller than 0.1 μm, the sedimentation mechanism of particle collection is negligible.

It was mentioned earlier that the impaction mechanism of collection in the foam scrubber may be assumed negligible. To evaluate the impaction one would need to analyze flow development conditions within the foam bubbles applying the Navier-Stokes equation. This is not an easy task.

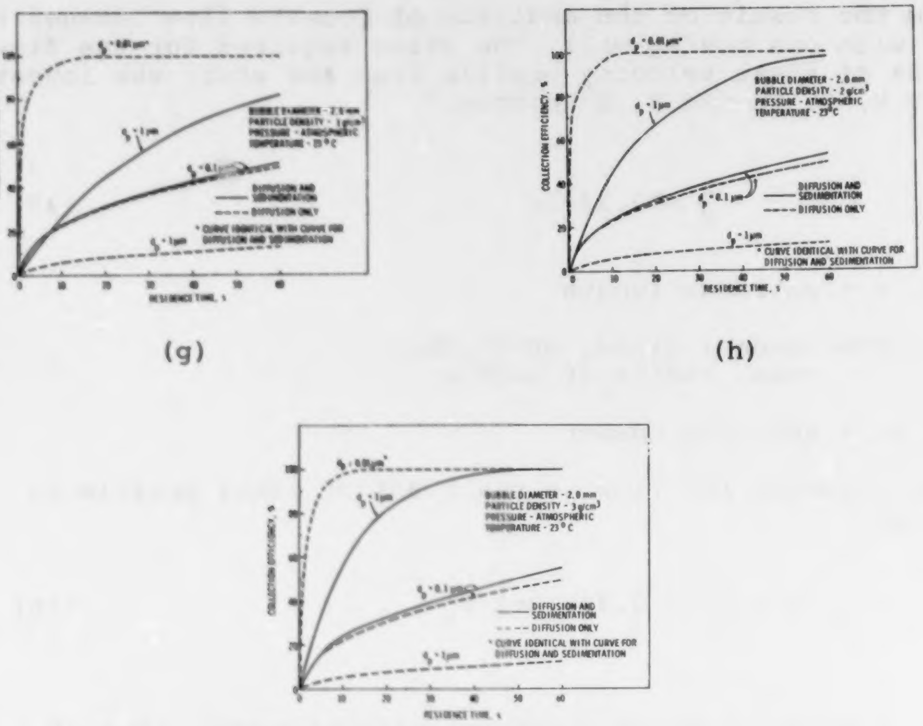


(a) (b)

(c) (d)

(e) (f)

Figure 4. Theoretical collection efficiency (cont'd next page)



(g) (h)

(i)

Figure 4. Theoretical collection efficiency (concluded)

However, as the first approximation, the order of magnitude of relaxation time for gas flow inside the bubbles to reach a steady state can be estimated from an analogous case of a gas bubble rising through a liquid, Figure 5. This will provide some indication of significance of the unsteady state on the overall collection process. Let us determine how fast the steady flow profile can be attained in the bubble illustrated in Figure 5.

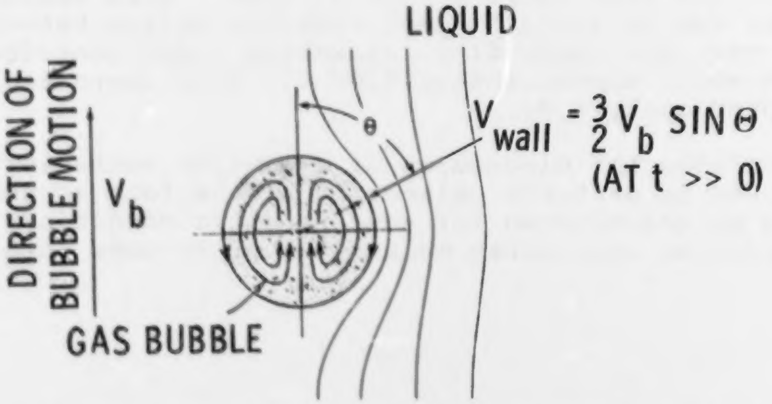

LIQUID
DIRECTION OF BUBBLE MOTION
V_b
θ
$V_{wall} = \frac{3}{2} V_b \sin\Theta$
(AT t >> 0)
GAS BUBBLE

Figure 5. Flow profile of a rising bubble

From the result of the analysis of Couette flow between the annulus with one moving wall, the value required for the flow to reach 95% of final velocity profile from the start was investigated by H. S. Yu and E. M Sparrow:[2]

$$\frac{x}{R} = 0.32\ \mathrm{Re} \qquad (15)$$

where x = entrance length

R = channel width, or in this case, radius of bubble

Re = Reynolds number

The time required for flow to reach 95% of final profile is therefore:

$$t = \frac{x}{V_b} = 0.32R(\mathrm{Re})/V_b \qquad (16)$$

For a bubble 1 mm in diameter, rate of bubble rise of 0.30 m/s, and a Reynolds number of 100, Equation 16 gives $t = 0.05$ second. That means the gas flow in a 1 mm bubble would reach the steady state almost immediately, say in less than 0.1 second. The motion of gas, however, would persist as long as the bubbles are rising through the liquid layer. By the same reasoning, as soon as the foam bubble is out of the liquid, the internal gas motion would subside immediately in the same relaxation time (<0.1 s) due to absence of surface film motion.

For in the case of a screen foam generator, initial disturbance of gas movement from the foam generation process would be expected to subside in the same short period of time. This would tend to indicate that due to insignificant relative motion between moving bubbles and gas, and short flow relaxation time, contribution of impaction would appear insignificant. This supports the assumption previously made.

This concludes the discussion of the major mechanisms that may be involved in particle collection by the foam scrubber. These should be reevaluated for some specific conditions of scrubber operation when other mechanisms might come into play.

EXPERIMENTAL EVALUATION OF FOAM SCRUBBER

Theoretical evaluations were most helpful in designing the experimental test unit shown in Figure 6. Many experiments were performed using this unit to learn more about foam generation, foam destruction, foam flow characteristics, foam stability, foam bubble size, and particle collection efficiency. It would be difficult to cover all of these tests in this presentation and the reader is referred to the references for additional details.[3,4]

Four surfactants were evaluated in our experiments. They are listed in Table 2. Most of the particle collection experiments were performed using Tergitol surfactant.

The foam destruction chamber was built to enable testing of various foam destruction techniques. Specifically, five foam destruction techniques were investigated. They included liquid spray, thermal destruction, ultrasonic destruction, compressed air spray, and high speed rotating disk.

The high speed rotating disk was found to be the most effective device for foam destruction because of low energy requirements and minimum deterioration of surfactant liquor qualities for a possible recycle. Consequently, this device was selected and used in particle collection efficiency evaluations.

Two types of aerosol particles were also tested. They were DOP (dioctyl phthalate) and polypropylene glycol. Relative to water the two particle types possess different wettability properties, the dioctyl phthalate being essentially nonwettable and the propylene glycol being wettable and miscible with water in all proportions. The experimental aerosols were generated by means of a condensation aerosol generator previously described.[5] Monitoring of particles was done by different particle counters including a nuclei condensation counter, an optical counter (Royco), and an electrical mobility analyzer. These instruments have a combined capability to measure aerosol particles in the size range from 0.003 μm to about 30 μm and were selected depending on the particle size range investigated at the time.

Collection efficiency measurements were made with different scrubber residence times. The scrubber residence times were varied by changing the scrubber length. At the same time the air flow velocity through the scrubber was maintained constant. This arrangement permitted performance of collection measurements with identical foam characteristics. The results of collection efficiency experiments are summarized in graphical form in Figures 7 and 8.

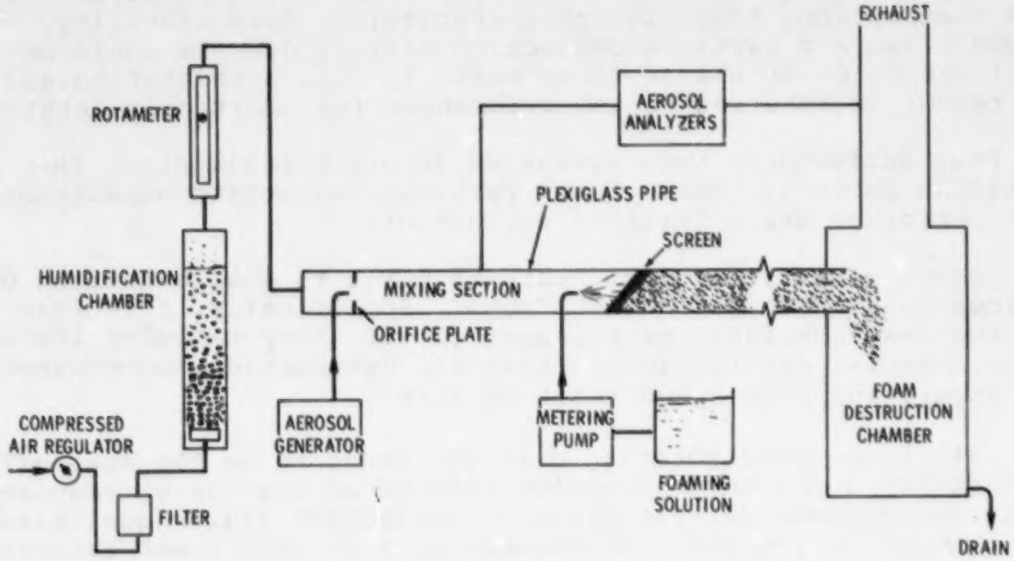

EXHAUST
ROTAMETER
AEROSOL ANALYZERS
PLEXIGLASS PIPE
SCREEN
HUMIDIFICATION CHAMBER
MIXING SECTION
ORIFICE PLATE
COMPRESSED AIR REGULATOR
AEROSOL GENERATOR
METERING PUMP
FOAM DESTRUCTION CHAMBER
FOAMING SOLUTION
FILTER
DRAIN

Figure 6. Experimental foam scrubber

Table 2. SURFACTANT TYPES

Surfactant	Manufacturer	Composition	Chemical type	Concentration, %	Comments
Tergitol TMN	Union Carbide Corp., Plastics Div.	Trimethyl nonyl polyethylene glycol ether	Nonionic	90	Wetting and rewetting agent
Sterox NM	Monsanto Co., Inorganic Chemicals Div.	Alkyl phenol ethylene oxide adduct	Nonionic	99.5	General purpose detergent (melting point 30-35°C)
Alkanol DW	E.I. du Pont de Nemours & Co., Inc.	Sodium alkylaryl sulfonate	Anionic	28	General use detergent; du Pont says it is in tight supply at the present time
Aerosol OT (75%)	American Cyanamid Co., Industrial Chemicals, Div.	Dioctyl ester of sodium sulfosuccinic acid	Anionic	75	Fruity pungent odor

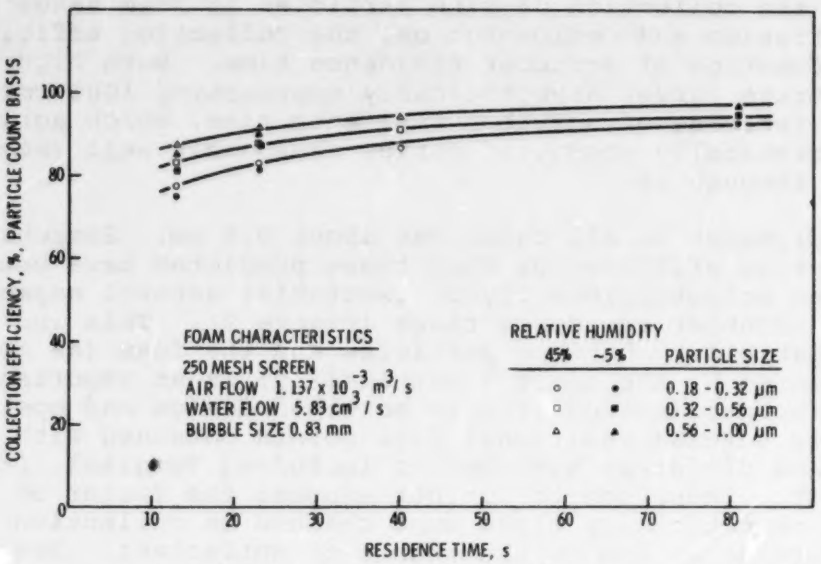


Figure 7. Foam scrubber collection efficiency of polypropylene glycol 425 using 2% Tergitol foam with rotating disk for foam destruction

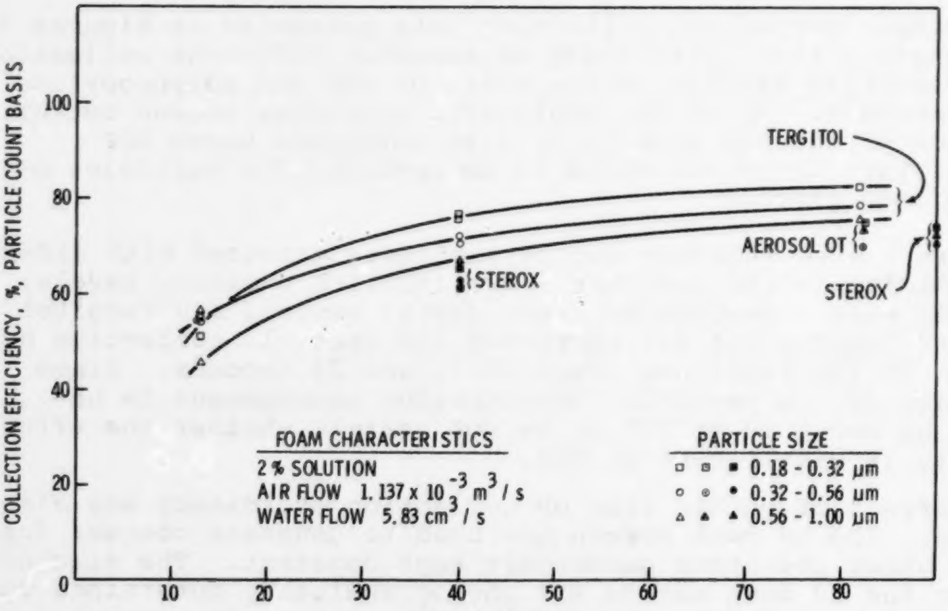


Figure 8. Collection of DOP aerosol using foam scrubbing with rotating disk for foam destruction

Because the collection of fine particles in foam takes place mainly by diffusion and sedimentation, the collection efficiency is a strong function of scrubber residence time. Both Figures 7 and 8 demonstrate curves asymptotically approaching 100% collection with an increase of scrubber residence time, which agrees with the theoretically predicted curves reasonably well (see Figures 4, a through i).

Bubble diameter in all cases was about 0.8 mm. Somewhat higher collection efficiencies than those predicted have been observed using polypropylene glycol (wettable) aerosol especially at the lower scrubber residence times (Figure 7). This indicates that the compatibility between particles and the foam (as for example suggested by the aerosol wettability) is an important factor to consider in optimization of foam scrubber design and operation. Figure 8 shows several additional data points obtained with the DOP aerosol and different surfactants including Tergitol, Sterox, and Aerosol OT. These points further support the factor of aerosol/foam compatibility since some changes in collection efficiency were noted due to the change of surfactant. Even though the collection efficiencies were found to be generally lower than those observed using Tergitol foam, proper surfactant selection might improve DOP collection and approach the efficiency that was observed for polypropylene glycol aerosol.

A comparison of the collection data presented in Figures 7 and 8 suggests that there might be somewhat different collection rate controlling factors in the cases of DOP and polypropylene glycol aerosols. While the small size particles showed better collection efficiency than large size particles using DOP aerosols, this trend was found to be reversed for particles of polypropylene glycol.

Figure 7 also includes two sets of data measured with different humidity in the scrubber. Two relative humidity levels, 5% and 45%, were investigated using glycol aerosol and Tergitol foam. More humid input air increased the particle collection by about 1 to 2% for residence times of 13 and 26 seconds. Since the accuracy of the particle concentration measurement is believed to be about 10 to 20% it is not certain whether the effect of humidity is significant or not.

The effect of bubble size on collection efficiency was also evaluated. The 60 mesh screen was used to generate coarser foam with the other operating parameters kept constant. The size of cells for the 60 mesh screen was photographically determined to be about 3.9 mm. The observed collection rates are presented in Figure 9. The collection rates are significantly less than the values determined for the small-celled foam (0.83 mm). The DOP aerosol showed a collection rate of about 40% with a particle size between 0.56 and 1.0 mm and 80 second residence time (75% collection was observed with the small-celled foam). The glycol achieved about 84% collection under similar conditions (98% with small-celled foam). Two data sets are also displayed using a large-celled 1% Alkanol foam. While this foam improved collection of DOP aerosol, the collection of polypropylene glycol aerosol was essentially unchanged.

The operating conditions which have been used in our experiments give an air (or foam) velocity in the scrubber of about 0.15 m/s (0.5 ft/s). It is postulated that a higher average velocity might influence the collection rate, especially at the point of foam generation where higher turbulence could enhance the effect of collection by impaction. Consequently, the velocity was increased by a factor of three to about 0.45 m/s (1.5 ft/s). At the same time the scrubber was lengthened to 5.5 m (18 ft) so that a 13 second residence time remained unchanged. The collection results for both DOP and the polyglycol were essentially the same as for the 13 second residence time at the lower air velocity.

This supports the theoretical evaluations presented earlier which indicated that the collection rates during the foam formation will be negligible and that the majority of particles will have to be collected by diffusion and sedimentation mechanisms.

There does, however, appear to be a point where collection efficiency is enhanced by capture of particles in the foam generation step. This occurs at high liquid-to-gas ratios causing the foam scrubber to behave similar to a wet scrubber. Figure 10 shows the effect of distilled water sprayed on the 250 mesh screen in relation to collection efficiency of DOP particles in the size range between 0.7 and 1.0 µm. No collection of particles is seen for liquid-to-gas ratios below 5,130 cm^3/m^3 (38.4 gal./1,000 ft^3). Thus, the collection efficiencies for foam scrubbing reported above primarily represent the mechanisms of diffusion and sedimentation.

Experience from bench scale operation was used to scale-up the foam scrubber to a small pilot unit with a capacity of 500 acfm. The objectives of this scale-up may be summarized as follows: (1) demonstrate the foam scrubber on a larger scale, (2) test it using an industrial aerosol particle, (3) gain experience in scale-up, and (4) verify a potential for multiple surfactant liquor recycles. It may be said that all these objectives were met with a considerable degree of success.

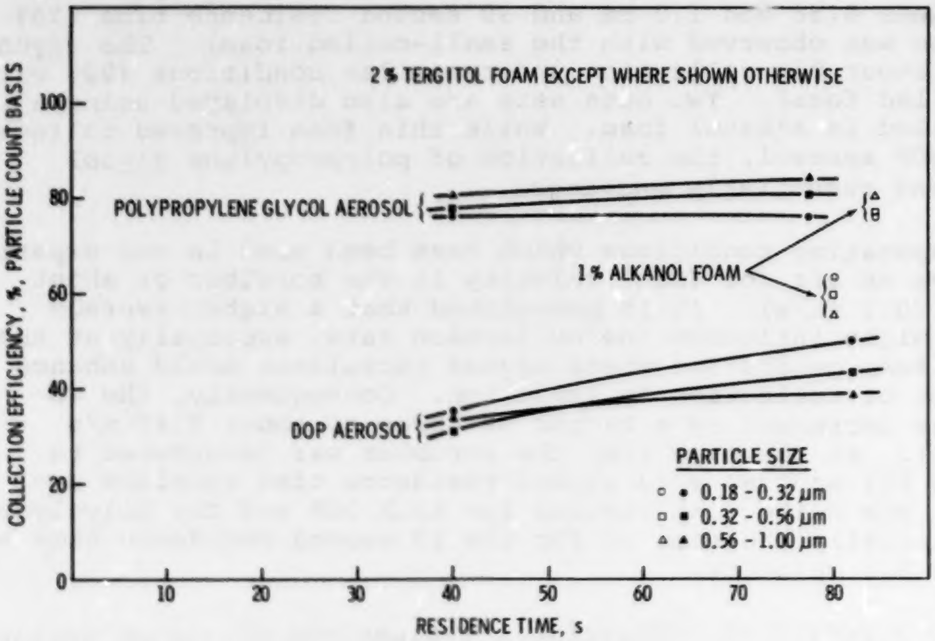


Figure 9. Foam scrubber collection efficiency of polypropylene glycol 425 and DOP using large-celled (3.9 mm) foam

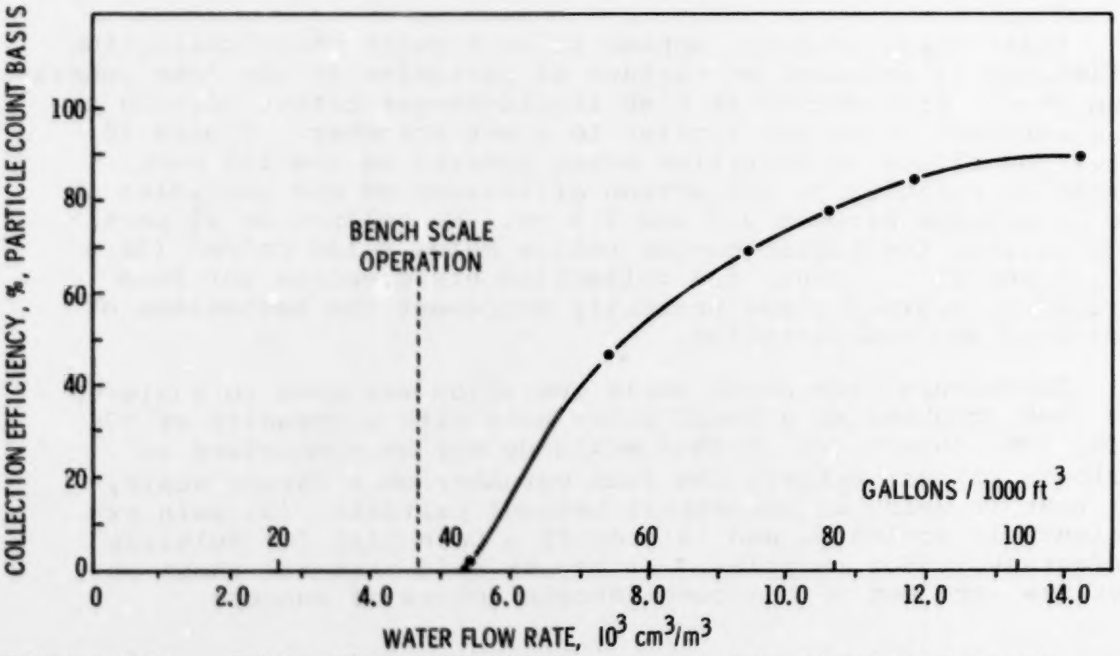


Figure 10. Collection efficiency values for distilled water sprayed on the screen versus water flow rate, DOP particle size: 0.7 - 1.0 μm

The scrubber was scaled-up and operated with no difficulties for several weeks. Collection efficiencies were measured using an industrial aerosol, a flyash obtained from a stoker coal-fired, power plant, Nucla, Colorado. The schematic of the small pilot scale foam scrubber is shown in Figure 11.

As shown in Figures 3 and 4, the particle collection efficiencies follow curves that steadily increase with time and asymptotically approach 100%. After a certain time, the gain in particle collection efficiency that would result with a further increase of residence time becomes rather negligible, and gets within the range of experimental error. To operate the pilot scrubber in this region would greatly reduce the possibilities for observing the effect of influential variables on the scrubber collection efficiency. It was decided that high scrubber collection efficiencies obtainable with longer scrubber residence times would be sacrificed for the purpose of having a more flexible pilot test facility, and the scrubber was thus designed for short residence time.

The scrubber residence time was actually measured and determined to be 20.5 seconds. Other operating conditions are listed below:

gas glow	0.260 m^3/s	(550 acfm)
liquid flow	6.81 x $10^{-4} m^3/s$	(10.8 gpm)
scrubber ΔP	1.84 kPa	(7.4 in. H_2O)

Collection efficiencies for submicron particles were measured and are illustrated in Figure 12. They ranged between 50.0% and 75% by count for a particle size range of 0.056 μm to 1.0 μm. Corresponding theoretical collection efficiencies for particles in the size range between 0.1 μm and 1.0 μm and for the residence time of about 20 seconds are 32% to 67% by count (see Figure 4h).

The data in Figure 12 are average collection efficiencies over the period of 20 hours. The surfactant concentration during the 20-hour experiment was kept at 0.25% by weight. This is a significant improvement over the bench-scale phase where about 2% surfactant solutions were needed to generate satisfactory foams. In addition, the surfactant liquor requirements were reduced from 5.17 x 10^{-3} m^3 per m^3 of gas (38.4 gpm per 1,000 cfm) to 2.64 x 10^{-3} m^3 per m^3 of gas (19.6 gpm per 1,000 cfm).

During the 20-hour run, the surfactant liquor was recirculated. As a consequence of the recirculation, the concentration of particles in the scrubbing liquor steadily increased to a level of 2 kg/m^3. No difficulties were observed in generating foam with the surfactant liquor containing this concentration of solids.

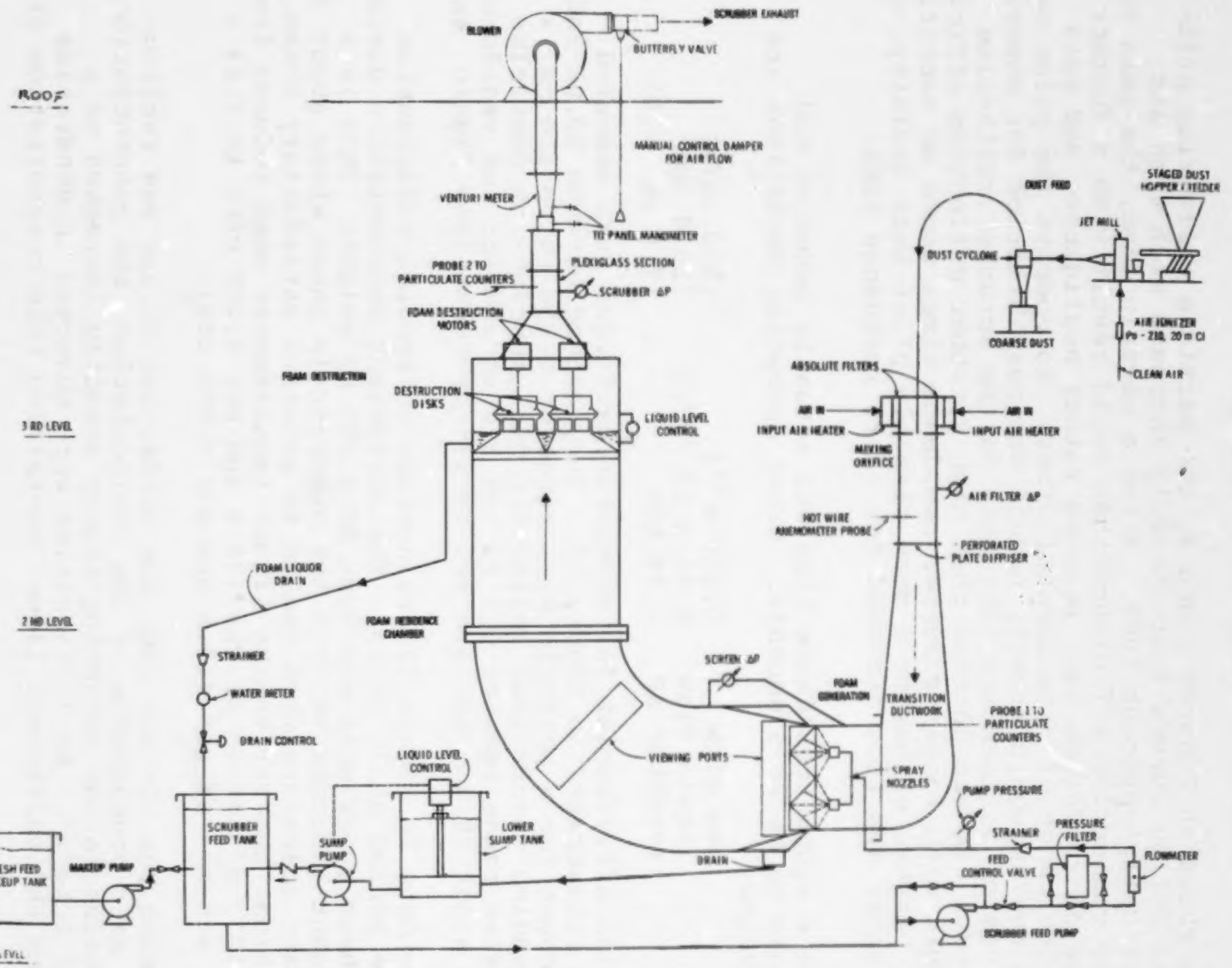

ROOF
3 RD LEVEL
2 ND LEVEL
GROUND LEVEL
BLOWER
SCRUBBER EXHAUST
BUTTERFLY VALVE
MANUAL CONTROL DAMPER FOR AIR FLOW
VENTURI METER
TO PANEL MANOMETER
PROBE 2 TO PARTICULATE COUNTERS
PLEXIGLASS SECTION
SCRUBBER ΔP
FOAM DESTRUCTION MOTORS
FOAM DESTRUCTION
DESTRUCTION DISKS
LIQUID LEVEL CONTROL
FOAM LIQUOR DRAIN
FOAM RESIDENCE CHAMBER
STRAINER
WATER METER
DRAIN CONTROL
LIQUID LEVEL CONTROL
LOWER SUMP TANK
SCRUBBER FEED TANK
SUMP PUMP
FRESH FEED MAKEUP TANK
MAKEUP PUMP
VIEWING PORTS
DRAIN
SCREEN ΔP
FOAM GENERATION
TRANSITION DUCTWORK
PROBE 1 TO PARTICULATE COUNTERS
SPRAY NOZZLES
PUMP PRESSURE
STRAINER
PRESSURE FILTER
FEED CONTROL VALVE
FLOWMETER
SCRUBBER FEED PUMP
DUST FEED
STAGED DUST HOPPER / FEEDER
JET MILL
DUST CYCLONE
COARSE DUST
AIR IONIZER Po - 210, 20 m Ci
CLEAN AIR
ABSOLUTE FILTERS
AIR IN
AIR IN
INPUT AIR HEATER
INPUT AIR HEATER
MIXING ORIFICE
AIR FILTER ΔP
HOT WIRE ANEMOMETER PROBE
PERFORATED PLATE DIFFUSER

Figure 11. Pilot-scale foam scrubber process layout.

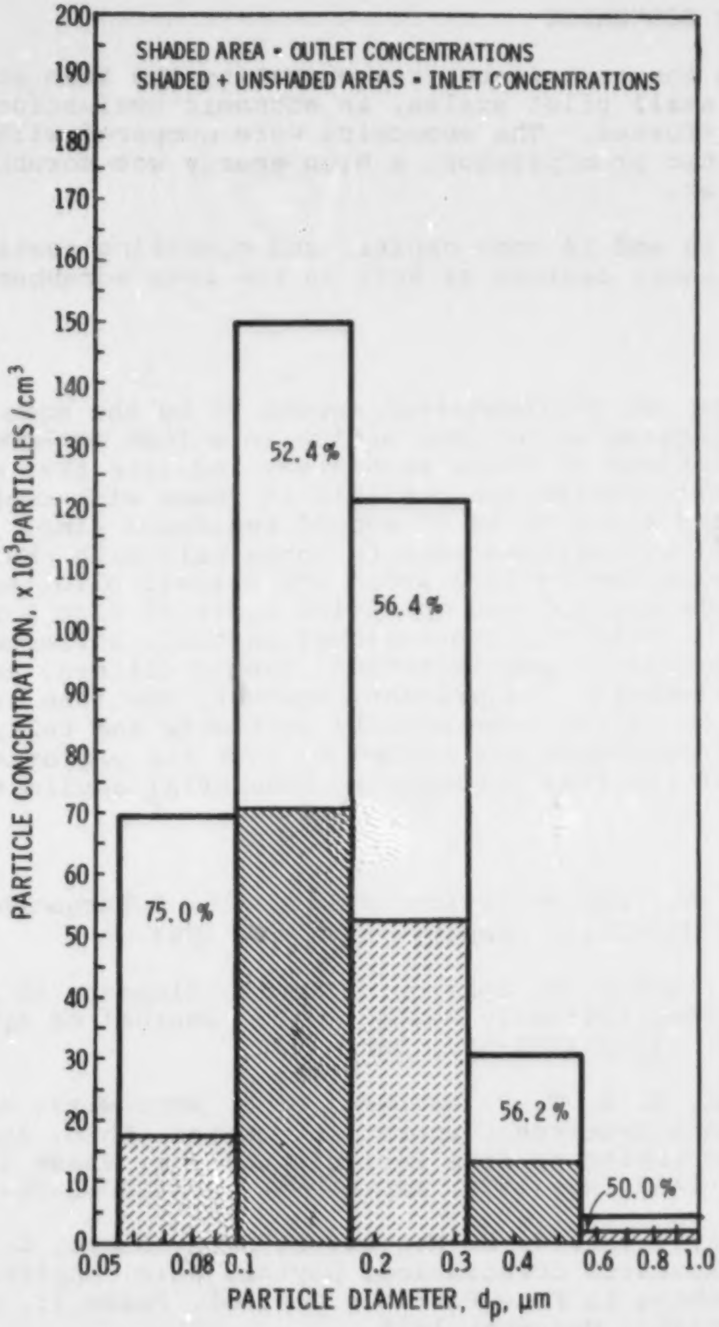


Figure 12. Average particle collection efficiency data (12 runs). Percent collection shown inside the unshaded area. Scrubber residence time 20.5 seconds.

FOAM SCRUBBER ECONOMICS

Based on the experience from operating the foam scrubber on bench and small pilot scales, an economic evaluation of this device was performed. The economics were compared with those for an electrostatic precipitator, a high energy wet scrubber, and a fabric filter.

Figures 13 and 14 show capital and operating costs for the three conventional devices as well as the foam scrubber.[5]

CONCLUSIONS

Diffusion and sedimentation appear to be the most important particle collection mechanisms active in a foam scrubber. Theoretical evaluations of these mechanisms indicate that over 90% collection efficiencies are possible in foams with bubbles 1 mm in diameter and about 40 to 60 second residence time. The results of theoretical evaluations seem to agree well with the results obtained by experiments in a bench and a small pilot scale foam scrubber. Both capital and operating costs of foam scrubbing are comparable with the conventional particle collectors including electrostatic precipitators, fabric filters, and high energy wet scrubbers. At present, however, the foam scrubber could hardly be called commercially available and fully proven. Additional developments are needed to test the performance and reliability of the foam scrubber in industrial applications.

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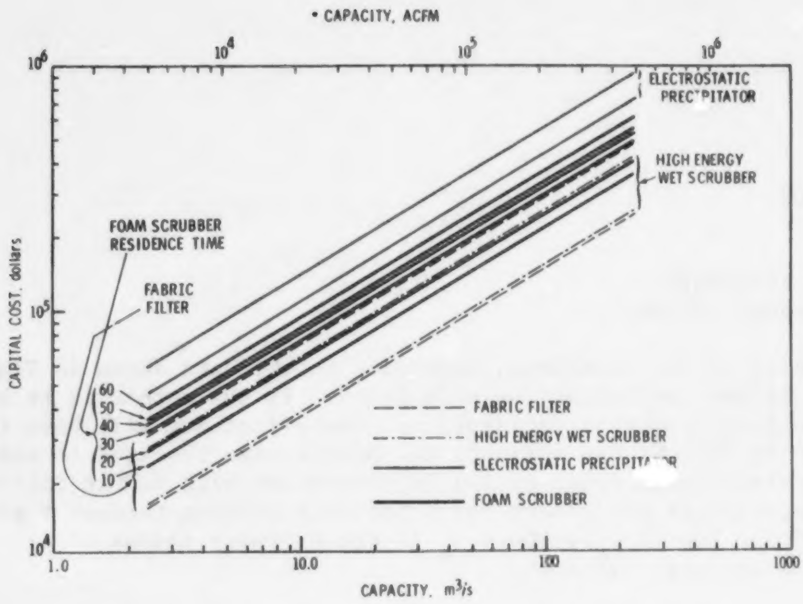

Figure 13. Capital cost for particulate control devices

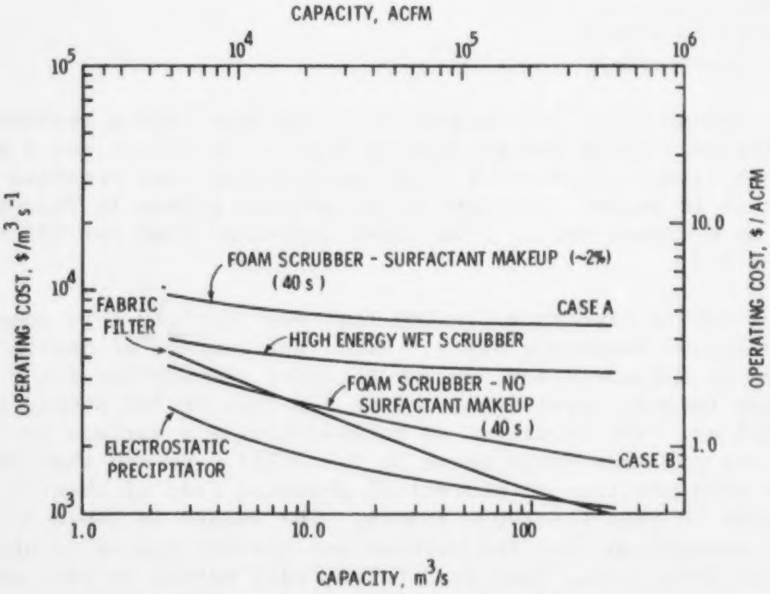

Figure 14 . Operating cost for particulate control devices

WRITTEN DISCUSSION

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The main problem of the technique, according to the data shown in Figure 12 of the paper, is the low collection efficiency. If the method is to be acceptable for pollution control applications, the efficiency will have to rise to the order of 99%. We can estimate the bubble size required to achieve this by using a relationship found at TRI in connection with fabric filtration. We have found that capture efficiency for a particle passing through a pore was a function of the log of r_e/s where r_e is the Brownian radius of a particle, and s is the pore diameter.

The Brownian radius is given by

$$r_e = \sqrt{\frac{kT}{3\eta\pi}} \cdot \sqrt{\frac{t}{r}} \cdot \sqrt{1+0.9\,\frac{l}{r}}$$

where η is the viscosity of air
l the mean free path in air
r the particle radius
t the residence time

Table I gives values of r_e for various particle sizes for a residence time of 10 seconds. Figure 1 shows values of r_e/s where s is 0.5, 1 and 2 mm, where s is now the bubble diameter. Capture efficiency values were obtained from Figure 4a,d and g at 10 second residence times and all points in Figure 1 seem to fall on a common curve. The curve indicates that for 99% efficiency, r_e/s must be about 0.3.

Table II gives bubble diameters (in μm) that will satisfy this requirement at 10, 20 and 100 second residence times. Since the capture of one-μm particles is aided by sedimentation, we may consider the bubbles for 0.5 μm particles to be the largest permissible and we see that at 100 second residence time, these are 1/4 mm. The formation of bubbles having a surface tension of 70 ergs/cm^2 requires pressure drops shown in Table III. We see that formation of 1/4 mm bubbles only requires a theoretical pressure drop of about 7 inches of water and is thus within tolerable limits. The values in Table III were calculated on the assumption that the bubbles are already joined to neighboring bubbles as they form, i.e., that each bubble wall serves as the common wall to two adjacent bubbles. If the bubbles are formed individually, the pressure drops in Table III must be doubled, since a bubble has two surfaces.

TABLE I

BROWNIAN RADIUS OF VARIOUS PARTICLES

Particle radius r (μm)	Brownian radius, r_e (μm)
1	16
0.5	23
0.1	66
0.05	114
0.01	482

TABLE II

BUBBLE DIAMETERS GIVING 99% CAPTURE EFFICIENCY

Particle radius (μm)	Residence time, seconds		
	10	20	100
1	51	72	160
0.5	78	110	250
0.1	220	311	700
0.05	380	540	1200
0.01	1600	2260	5000

TABLE III

PRESSURE DROP REQUIRED FOR FORMATION OF BUBBLES OF 70 ERGS/CM^2

Bubble Diameter (mm)	ΔP (mm H_2O)
1	42
0.5	84
0.25	168

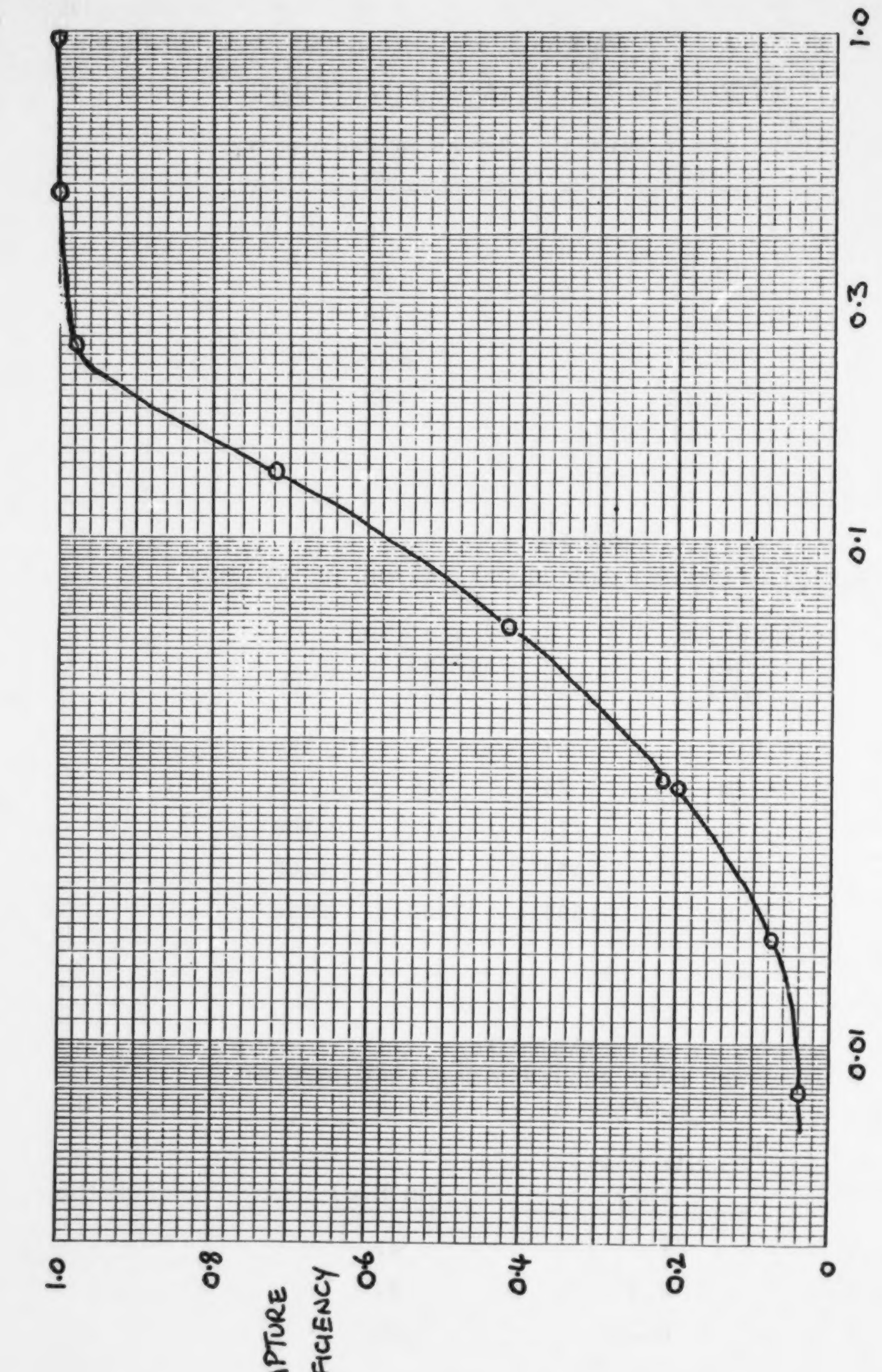

Fig. 1

The paper makes no mention of the effects of temperatures which must be limited to somewhere short of 100°C, and presumably limits application of the technique.

OPEN DISCUSSION

Cooper: I was struck with the lack of time dependence of the low collection efficiency values, for instance, in the DOP aerosols,page 16. One of the things being suggested is that there is the occurrence of another generation mechanism. We have the formation of very fine sodium chloride particles by the breaking of the bubbles. Maybe the device is better than we think because when the bubbles break, they are putting some of the material back into the air. I was wondering if you had a comment on that.

Ctvrtnicek: You mentioned that the collection efficiencies were high on the 500 cfm unit. I tried to explain that the unit was purposefully designed to have low collection efficiencies, that we wanted to be on that steep end of the curve. Our results were somewhat better than the theoretical collection efficiencies. We extended the time for deposition and still could not reach 900 gauss collection efficiency. Whatever we achieved was not achieved through any extraordinary effort to optimize the device. We were primarily after a demonstration of the model. We never optimized bubble generation, or the geometric structure. For example, in the bench scale, we were generating foam destruction on the order of 10^6 particles per cubic centimeter. When we went to the pilot scale, we cut that down by a factor of 10. I believe that it is very likely that, by proper geometry of the disc and impaction surfaces,this can be reduced even further. The same goes for the foam generation. We really did not try to optimize it at all. Once we got a decent foam we could use to make experiments, we just performed the experiments. There were no optimization efforts at all.

Lawson: You mentioned that the collection efficiency might be enhanced if you had smaller bubbles. Unfortunately, one of the undesirable effects, with the understanding that there has been no real effort to optimize the situation, is that if you go to smaller bubbles, you probably need even more surfactant to get the total bubble volume that is required. That is a possibility because the wall surface of the bubble will probably not go down nearly as fast as the size of the bubble. You will end up making smaller bubbles requiring even more surfactant. So surfactant recovery will really be important in that case.

Ctvrtnicek: That is a good point. However, when you can believe in recycles, surfactant really does become a factor. In operating this unit we used semi-horizontal and vertical modes. I believed that the device worked much better

in the vertical mode because you were continuously getting a drainage of the foam. As the foam becomes drier, it is easier to destroy. You can reduce the energies for destruction and make it even more effective.

"THE ANOMALOUS BEHAVIOR OF ASBESTOS FIBERS"

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INTRODUCTION

Asbestos is a term used to include a number of fibrous silcate minerals including amosite, crocidolite, chrysotile, tremolite, and anthophylite. Because of its inflammability it is widely used for insulation in clothing and construction, for brake-linings, and as a binder with cement and asphalt.

Recently there has been evidence showing a 50-300% increase in cancers of the digestive system among insulation workers [1]. A significant increase of lung cancers [2] especially mesothelioma among workers in the asbestos mining has been found. In this case a significant synergistic effect was demonstrated for smokers. Finally, an examination of cancerous tissue indicated the pressence [3] of longer fibers (fiber length > 5μ); although no similar studies have been carried out for healthy tissue.

This latter effect provides some insight into the reasons for increased interest into asbestos fibers. Probably the major emphasis to date has been in the detection of fibers. First the concentrations are very low, in the order of 10^5 to 10^6 particles/liter for rainwater. This corresponds to a weight fraction of $10^7 - 10^6$.

Typical concentrations of fibers in air at rural locations are of the order of 10^3 fibers/m^3. Such concentrations require a lengthy sampling time which has so far precluded automated instrumentation [5]. Furthermore, as is discussed, below only a fraction of the fibers can be asbestos. In recent years the separation of fibers has become recognized as an important problem. For example, are only long fibers carcinogenic? The point is important in that the general population is typically exposed only to short fibers whereas industrial workers are exposed to a wide spectrum of particle sizes. There is, also, the need to separate fibers from spherical particles in industrial regions. Typically fibers provide only a small fraction of the total fibers.

One point that should be emphasized is the difference between the mineral fibers. Amosite and crocidolite are rod-like, they appear rigid, and their shape can be approximated by a prolate spheroid. Chrysotile which is the most widely used fiber consists of hollow, bent tubes which appear to be flexible. Anthophylite appears to be mostly plate-like resembling shingles. For most measurements amosite or crocidolite are recommended.

GENERAL PROPERTIES

Before the specific properties of non-spherical particles are discussed, it is well to outline the steps required for a systematic development of spherical particles. Condensation and evaporation are, of course, less important than for spherical particles, since non-spherical particles are solids. Similarly the problem of coagulation has been less studied, primarily because of the much more complicated expressions and the stochastic nature of the collision process. Other than these differences the problems are similar. First, the aerosols must be generated. The question is what is the analog of a monodisperse distribution of spherical particles. Next the particles must be measured. The methods of measuring non-spherical particles by microscopy, by aerodynamic behavior and by optical methods are related to those of spherical particles. In this lecture, emphasis will be focused primarily on the aerodynamic behavior because it is the most thoroughly investigated. Finally there are the separation processes - electrostatic precipitation, filtration, gravitational settling in an elutriator all of which are just beginning to be studied.

GENERATION

The ideal generator would produce a uniform, monodisperse cloud of fibers with a known and controllable charge distribution. A monodisperse fiber would be one having the same shape and the same dimensions. For the approximation to this fiber one would have an amosite fiber which can be approximated by a prolate spheroid. The monodisperse distribution would then be amosite fibers having the same length and diameter. Unfortunately we do not have the analog of latex particles or a Sinclair-La Mer generator for non-spherical particles.

Many of the current asbestos generators which work by scraping a packed cake of ground asbestos fibers are completely unacceptable for experimental studies of the properties of fibers in that the distribution is very broad consisting primarily of aggregate rather than the necessary individual fibers.

The most successful generator [6] currently available uses a vibrating bed with the flow rates through the bed at sub-fluidization velocities. The asbestos fibers are ground and dried prior to placing in the generator. Mixing the asbestos with small metal spheres (Nickel) proved unsuccessful in that the fibers produced by the generator were coated with a thin layer of nickel dust. The generator described above has been proven to give a narrow particle size distribution with the diameter of the fibers ranging from 0.2-0.4μ and the lengths from 1 to 10μ. The mean value of the aspect ratio (the ratio of major to minor axis) was approximately 10.

The generator can produce fibers in the concentration 10^2 - 10^4 fibers/cm^3. Examination of the fibers indicate that approximately 15% of the weight of the fibers is in agglomerates but approximately 85% of the weight is in individual fibers. The particles are mostly shorter than 5μ and are relatively highly charged in comparison to spherical particles.

It has been suggested that non-spherical particles can be generated using solutions which yield non-spherical crystals on evaporation [7]. Also, certain types of spores are oblong and could be used to model non-spherical particles. Neither of these suggestions have been successfully developed.

DETECTION

The only currently reliable method of determining the size and number of asbestos particles is electron microscopy. The key points that have been raised regarding microscopy of non-spherical particles can be summarized as follows:

1. Not all fiberous particles are asbestos.

2. Electron microscopy gives only a 2 dimensional projection. In order to unambiguously characterize the fiber it should be examined from three dimensions.

3. Nuclepore filters are the best current technology for collecting a sample.

The first result was demonstrated by Spurny and co-workers in their measurement of air in the mountains near the source of the Ruhr. They examined the elemental composition of each apparent fiber (particles with an aspect ratio > 3). Any fiber which contained Fe, Si, and Mg was classified as a possible asbestos fibers. These experiments showed that no more than 5% of the sampled particles could be asbestos. Similar results were found for air sampled from Münster a city of 200,000 without heavy industry.

In analyzing asbestos fibers, the prevailing assumption is that fibers lie flat on the surface, so that the particle dimensions can be obtained directly from the projection. Doubt has been raised regarding this point, and studies with spores ($1 < B < 3$) have shown some particles aligned with the surface. Because of the non-random orientation of particles, Cauchy's theorem cannot be applied to obtain a particle size and the particle must be viewed from three angles. The work required is painstaking and tedious. The hypothesis of the fibers lying flat has not been adequately tested.

Three methods have been examined for collecting samples - filtration through membrane or nuclepore filters, electrostatic precipitation, and thermal precipitation. The results of a comparative study are discussed extensively in a recent review article [5]. Two conclusions were drawn from the article - nuclepore filters are superior to membrane filters in the preparation of the sample and the filters require considerably less time per experiment than do the thermal or electrostatic precipitator.

AERODYNAMIC DIAMETER

Perhaps, the most extensively studied aspect of non-spherical particles is their behavior in free fall. In comparing experimental measurements with theoretical calculations - use is made of the implied assumption that the particles may be treated as prolate or oblate spheroids.

In discussing results for non-spherical particles it is necessary to define the aerodynamic diameter and the dynamic shape factor for a non-spherical particle.

The aerodynamic diameter D_{ae} of a particle is defined as a sphere of unit density having the same settling velocity as the particle at low Reynolds number. The aerodynamic diameter is related to the volume equivalent diameter D_e and the dynamic shape factor K by equating the settling velocities giving:

$$V_s = \frac{1}{18}\,\frac{\rho_o D_{ae}^2}{\eta F(D_{ae}, V_s)} = \frac{1}{18}\,\frac{\rho}{\eta}\,\frac{D_e^2\ K}{F(D_e, V_s)}$$

where the volume equivalent diameter is the diameter of a hypothetical sphere having the same volume as the particle. The function $F(D, V_s)$ is related to the drag coefficient (C_D) and Reynolds number (Re) by

$$F(D, V) = C_D \frac{Re}{24}$$

For the important limiting case of continuum mechanics and creeping flow the aerodynamic diameter is related to the equivalent diameter by

$$K = \frac{\rho}{\rho_o}\,\frac{D_e^2}{D_{ae}^2}$$

For spheroids and cylinders, the theoretical shape factors have been found for motion parallel to the polar axis K_{11} and perpendicular to the polar axis K_{22}. For a prolate spheroid these results are

$$K_{11} = \frac{4}{3}\,\frac{(B^2-1)\ B^{-1/3}}{\frac{2B^2-1}{\sqrt{B^2-1}}\ Ln(B+\sqrt{B^2-1}) - B} \qquad K_{22} = \frac{8}{3}\,\frac{(B^2-1)\ B^{-1/3}}{\frac{2B^2-1}{\sqrt{B^2-1}}\ Ln(B+\sqrt{B^2-1}) + B}$$

where B is the aspect ratio (the ratio of major to minor axis). Similar expressions exist for cylinders and oblate spheroids [8].

Several key points should be stressed regarding the shape factors for non-spherical particles. First the shape factors are a function of the aspect ratio alone. Secondly, the ratio of K_{22}/K_{11} is bounded between 1 and 2. The function $\frac{K_{22}}{K_{11}}$ is a slowly increasing function of B.

The effective or experimental value of the shape factor can be calculated by

$$K = \int_{0}^{2\pi} \delta\theta \; [K_{11} + K_{22}] \; f(\theta) \; \sin\theta$$

where f(θ) is the angular distribution of the particle orientation. It is precisely the difficulties with the form of f(θ) which is the subject of much uncertainty and which accounts for the curious behavior of non-spherical particles that have been observed experimentally. These effects can be summarized in the table below.

Table 1

1. Random-Orientation Assumption
2. Gans Diffusion Solution
    a. Gallily-Cohen Simulations
    b. Comparison with Random-Orientation Assumption
3. Collection Efficiency Measurements
    a. Spiral Centrifuge
    b. Inertial Impactor
    c. Filters
    d. Elutriators

The basic result of all experimental measurements have been that the particles tend to acquire a preferential orientation, usually in the optimum position for penetration, in any flow field where there are inhomogenieties in the flow field.

For the special case of free-fall of particles in a stationary medium, the shape factor can be computed if the functional form of f(θ) is known. By suitable experiments, experimental value of the aerodynamic diameter can be measured and compared with theory. Assuming creeping flow and a continuum, one has

$$D_{ae} = \left[\frac{18\eta V_s}{\rho_o} \cos\alpha \right]^{1/2}$$

The product $V_s \cos\alpha$ is the vertical component of the settling velocity where α is the angle of descent of particles. As is known, non-spherical particles do not descend vertically, but at an angle α [9] which can be related to the ratio of the shape factors for movement parallel and perpendicular to the major axis and the angle of orientation θ. The angle α is defined by

$$\tan\alpha = \frac{[(K_{22}/K_{11})-1]\tan\theta}{(K_{22}/K_{11}+\tan^2\theta)}$$

In a recent note [10], comparisons were made of calculations for the aerodynamic diameter using the assumption of random orientation, the assumption of an averaged orientation, and a stochastic simulation where the particle was subject to frequent changes in orientation with the probability chosen to correspond to Gans solution [12]. It was found that the average value of the aerodynamic diameters were similar for all three calculations, but that the scatter of measurements or the standard deviation of the aerodynamic diameter was less for the numerical simulations than found for the assumption of random orientation. It should be stressed that the two calculations are based upon different assumptions, and that the simulation gives a lesser degree of randomness manifested by a lower standard deviation than does the assumption of random orientation. The discrepancy indicated here emphasizes the need for experimental measurements to check the assumption of random orientation (which is the prevailing assumption) and the need for looking at variances of stochastic processes rather than mean quantities.

Quanitative experiments in which the aerodynamic properties of non-spherical particles have been examined include sedimentation experiments in silicone oil, experiments with the spiral centrifuge, experiments with an inertial impactor, and experiments with nuclepore filters.

The spiral centrifuge is widely used for particle separation and analysis of polydisperse aerosols. The particles are removed by centripetal force as a function of their aerodynamic diameters. Particles having a smaller aerodynamic diameter are deposited on a metal strip lining the walls of the spiral nearer the gas inlet then larger particles. The advantage of the instrument is that there are distinct regions of deposition depending on the aerodynamic diameters.

Experimental measurements (primarily by Stöber [13] and coworkers) have found the following important results:

1. The aerodynamic behavior of clumps of monodisperse latex spheres differ depending on whether they are ball-like or chain-like clusters. Specifically

$D_{ae} \propto N^{1/6}$ for chain aggregates

and $D_{ae} \propto N^{1/3}$ for clusters.

2. Experiments with aerosols generated from an exploding wire in which the particles showed a high degree of agglomeration could be correlated with similar expressions as the latex particles [14]. Two separate regions were found with R_a for the lower values of N correlating according to the chain-like clusters

$$D_{ae} \propto N^{1/6}$$

3. Experiments with asbestos fibers (amosite and crocidolite which both provide rod-like fibers) showed that the aerodynamic diameter could be correlated with expressions of the form [13]

$$D_{ae} \simeq a_1 D_1 (L/D)^b$$

where D is the diameter of the fiber. a was found to be approximately 2 and the exponent b ranged between 0.1 and 0.2. It should be noted that there was considerable scatter in the data and that the L/D ratio used in this study was approximately 10.

In summary the spiral centrifuge can be used to separate non-spherical particles, however the separation is strongly related to the diameter and a weak function of the aspect ratio indicating that other methods will probably be necessary in order to determine the length. It should be noted that the deposition length in the spiral centrifuge is a calibrated quantity.

OPTICAL METHODS

The optical properties of fibers are not only effected by the physical properties of the fiber but the angle of orientation as well. Although the problem of back-scattering from an arbitrary prolate spheroid has been solved for a conducting spheroid at direct incidence [15], the problem for a general orientation has not been. Further as pointed out in the previous section the orientation of the fibers are generally not known.

Several investigators have measured the fiber properties with a Royco. Early experiments reported good agreement with microscopic [16] examination for large fibers, but recent experiments show that the distribution of the "apparent [17] optical diameter" is significantly broader than the distribution of the diameter, length, or aerodynamic diameter determined from measurements of the individual particles. The location of the maximum for the aerodynamic diameter and the "optical" diameter almost overlapped.

By modifying the Royco so that higher velocities could be obtained Gentry and Spurny demonstrated an apparent narrowing of the distribution. This effect was attributed to the particles acquiring a less random orientation. Their experiments suggest that forcing the particles to acquire a preferred orientation may be the only method for measuring the particle distribution by light scattering.

Three methods have been suggested for imposing a preferred orientation on the particles. These methods are electrostatic forces, magnetic fields, and by increasing the flow rate.

Generally speaking, all the basic problems - the rate unipolar charging of the particles, the distribution of charges on the particles, the mobility of particles in an electric field and the charge distribution are unsolved. Preliminary measurements of doublets of latex particles [18] in an electrical field show an alignment with the particles aligned parallel to the field. Qualitative experiments indicate that fibers tend to be more heavily charged than spherical particles.

The use of magnetic field to align particles has been suggested by Timbrell [19] as a means of particle separation. The method has been tested in quiescent liquids. However, the use of the method for particles in air has not been tested as yet. There are several drawbacks in the proposed method which can be indicated as follows:

1. The fibers may take orientations parallel or perpendicular to the field, including both orientations.

2. The preferred orientation can not be predicted in the absence of experiment, and

3. When fibers are cleaved along the major axis, the preferred orientation may be opposite of the parent fiber.

It should be pointed out that for asbestos fibers no treatment is necessary and that it has been shown that non-magnetic fibers can be treated so that they can be aligned in the field.

There is little doubt that the presence of a small gradients can induce alignment in fibers. The results are striking, as demonstrated by the following set of photographs taken for a series of nuclepore filters in series. The filters range through the sizes 8.0μ to 0.2μ. The length and diameters of the fibers were measured.

What is striking about this measurement is that the particles deposited on the last stage penetrated a filter with openings of 0.6μ diameter. Since the average dimension of the fibers was a length of 2.3μ and a diameter of 0.4μ, only if the fibers were aligned with the gas stream lines could they penetrate to the backing filter.

Similar results have been found for a cascade impactor where almost 80% (by mass) of the particles penetrate to the backing filter although measurements of the aerodynamic diameter (calculated from electron micrographs of the fibers) indicate most of the particles should be deposited on the 4th or 5th stage.

SUMMARY

The characterization of asbestos fibers (or non-spherical particles) provide a number of significant differences to the characterization of spherical particles.

For efficiency measurements of spherical particles it is necessary to measure only the relative intensities of monodisperse particles. For fibers there are no monodisperse particles, and the amount of light scattered depends on orientation as well as the length and diameter.

For spherical particles the distribution can commonly be characterized by a log-normal distribution. For non-spherical particles there are additional dimensions that must be accounted for in the distribution. In addition average parameters such as the optical diameter can not be characterized by a simple distribution.

There, however, have been significant discoveries that indicate the directions of the future. These include

(i) The development of a generator capable of giving reproducible distributions of individual fibers,

(ii) The demonstration that only a small fraction of fibers in ambient air are asbestos,

(iii) The demonstration that the tendency of fibers to align with the gas stream lines in the presence of small gradients is sufficiently great that inertial impactors are useless, and

(iv) The proof that electrical, magnetic, or differences in flow gradients can impose a preferred orientation on the fibers.

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OPEN DISCUSSION

Gentry: The reason I presented more problems than solutions, is because there are more problems than solutions. Secondly, $\frac{L}{D} > 3$ and should be 6.

Friedlander: At the bottom of page 9, about the use of magnetic fields, no statement is made about what the magnetic and dielectric fields are.

Gentry: What has been suggested here is using the property that the particles will fall at different angles as a method of separation. A series of studies have been carried out that investigated this type of thing. First by dropping particles without a magnetic field and then with a magnetic field to align them. The particles were aligned in aqueous solutions which solidified. We looked at the solidified mass and found that the particles aligned either parallel or perpendicular to the field, depending on which asbestos was used. It was not merely a matter of whether the magnetic field was on or off. For example, by using permalite from Uganda and permalite from Uganda and from South Africa, there were different preferential orientations.

Oder: I have been involved in a lot of mineralogical studies with asbestos, and generally speaking, asbestos is associated with iron ore processing. The magnetic properties of asbestos are usually associated with the impurities attached to the surface. When I looked at your slides, the thing that struck me was that the crystals seemed to be very clean. Nothing was on the surface at all.

Gentry: Let me point out that the asbestos, which is pure (at a commercial grade) to start with, is not of the same magnitude of particle generation as that which is usually dealt with. Our generator output is about 13 milligrams to .13 milligrams per cubic meter. This is probably why the particles are so clean.

Oder: Since you have cleaned off the fibers, the fibers are not as they would occur in nature, but as attached to magnetic particles. The magnetism left over would probably be diamagnetism for fibers that would have very high substitutional iron or paramagnetic depending on the iron composition.

Friedlander: I would go further than that and say that just by taking a chemical analysis, you can not tell. An example is that the alpha and gamma phases of the iron would be quite different. Or maybe for iron oxide, one case would be parallel and the other ferromagnetic. So just knowing the composition does not tell you the answer either.

Gentry: If you use chrysotile, it will cleave along the major axis. The fibrils will then have a different magnetic orientation.

Friedlander: The paper says: "The presence of small gradients will produce alignment of fibers." What gradients are you talking about? Flow field? Magnetic field?

Gentry: Gradients of flow field.

Billings: Isn't asbestos of several different types? Aren't some quartz and some serpentine or magnesium?

Gentry: In our study, we defined three types of asbestos. They are chrysotile, crocidolite and amosite. Other people have carried out studies with tremolite and other asbestos types, but all basically consist of iron, magnesium and silicon. Silicon dioxide is not considered asbestos.

Oder: Isn't there a dependence between biological activity and fiber length?

Gentry: It is speculated that way and studies have been done. It has always been associated with longer particles. It is an interesting point now that carcinogens are important. People have suggested that only particles over 5 microns have any carcinogenic effect. The size dependence is determined by autopsy. If you look at a lung with a lesion, you will only find particles of over 5 microns.

Billings: I would like to comment on your measurement technique. There is a tendency for particles to bounce on the impactor stages and if you are not using some type of grease, there is a high probability that these particles just bounced all the way through. Once they bounce, they will go all the way to the back-up filter, because you do not have an efficiency going from zero to 100%. If goes from zero up to 90% then goes back down.

Gentry: I am fairly certain that for the most part there was a backing filter of petroleum jelly on each stage. This was used as a substrate for collecting particles and acts somewhat as an impactor.

The 90 to 95% of the particles that look like asbestos were predominately calcium sulfate. The analysis of chemicals was obtained by x-ray phoresence.

Löffler: Do the particles coming out of the fluidized bed carry any charge? What is their behavior?

Gentry: They are charged. We put the particles through a decharger to study them.

HIGH TEMPERATURE FILTRATION - TECHNICAL PROSPECTS AND AS MEANS FOR CHECK ON THEORIES.

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ABSTRACT

Adding temperature as a variable to the classic test parameters (particle size and face velocity) demonstrated that the flow field in fibrous filters,even at low fiber Reynold's number, shows departures from laminarity.Two phenomena might be involved: a. unsteady flow at capillary entrances and branchings and b. secondary flow.

These turbulence-like phenomena improve the filter performance in the sense of lowering pressure drop and increasing efficiency.

It is surmised that increasing the turbulence within the mat could result in improved industrial filters.

O.INTRODUCTION

Filtration through fibrous media at elevated temperatures is a well established practice in the cement and other industries. From the environmental point of view this process has the advantage that buoyancy of the hot gases is preserved and provides better dispersal.Nuclear technology also uses high temperature filtration.Considering its actual and potential importance,it is amazing that but so few papers were published about filtration at higher temperatures : First et al.(1),Billings et al.(2),Spaite et al.(3),Thring and Strauss (4),Pich and Binek (5),Strauss and Lancaster (6),Dyment (7).But when you get down to experimentation and notice that tests which normally take half an hour need a whole day or more at high temperature than the amazement stops. Some of the difficulties encountered are: the nature of the aerosol and its dispersal; problems in attaining thermal equilibrium within the apparatus; difficulties in attaining gas-tightness; delay of cooling of the apparatus to enable the change of the filter in view of weighing...).It goes without saying,that when "high temperature" is not 500°C for research purposes but just 100°C for technical assessment as described in Sect. 3.2, the experimental obstacles become less forbidding.

Besides its technical importance, tests at elevated temperatures allow deeper comprehension of the real behavior of fibrous filters.Given the well known inaccuracies of the filter testing, it is not surprising to see many a theory confirmed or at least not being contradicted by experiments made in the three-dimensional efficiency-particle size-face velocity space.But with one more dimension added,namely temperature,the challenge on the model-concept becomes harder.Dimensional analysis with respect to temperature gives a hitherto unsuspected insight in the physics of filtration.Discrepancies which seemed previously incidental, become apparent and this way our understanding about what happens might be improved.

Two aspects of the behavior of filters at elevated temperatures are of concern here : their pressure drop and their efficacity.The subject has been reviewed more or less recently in a very comrehensive way by several authors : Chen (8),Pich (9), Spurny (10) and Davies (11).Thus we assume that the theory of aerosol filtration by fibrous filters as well as the basic terminology are well known and therefore no details will be given here.

1.PRESSURE DROP AT ELEVATED TEMPERATURES

There are four lines of basic approach for the understanding of the pressure drop of filters:

a.The channel theory, b.the drag theory, c.the fiber drag and ineraction (e.g. the Kuwabara(12) flow field) and d.the geometrical theory of pore structure (13).

Even the hydrodynamically most elaborate of these theories as those of Happel (14), Fuchs and Stechkina (15) or Spielman and Goren (16) and which at room temperature are experimentally well verified,show one common feature : the pressure drop is directly proportinal to the dynamic viscosity of the carrier gas.The reason is the more or less implicit use of Stokes' lamimarity criterion,i.e. that the streamlines have up and downstream symmetry.

Our experiments show against all theoretical expectation that at variable temperature the pressure drop does not follow exactly the increase of the gas viscosity: Figs 1 and 2 show that the pressure drop of the media tested either in air or in argon increases less steeply than does the dynamic viscosity.For all experimental detail,see Quetier (17).

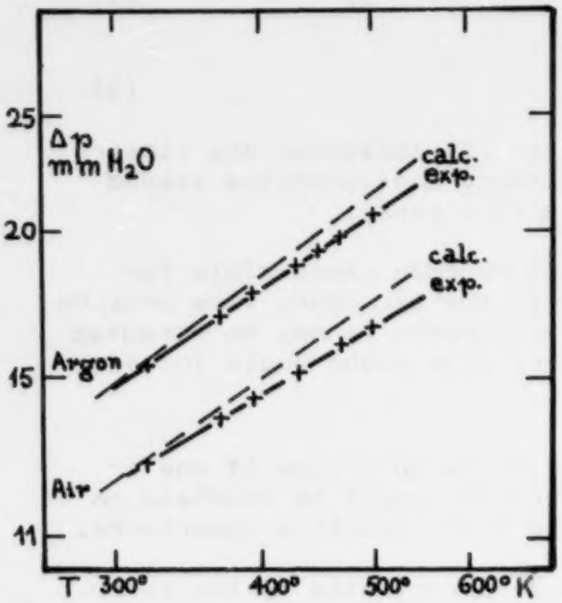


Fig. (1) - Increase of the pressure drop with temperature of a sintered glass filter,average pore diameter 40 - 90 micron,at constant face velocity,in air and argon.

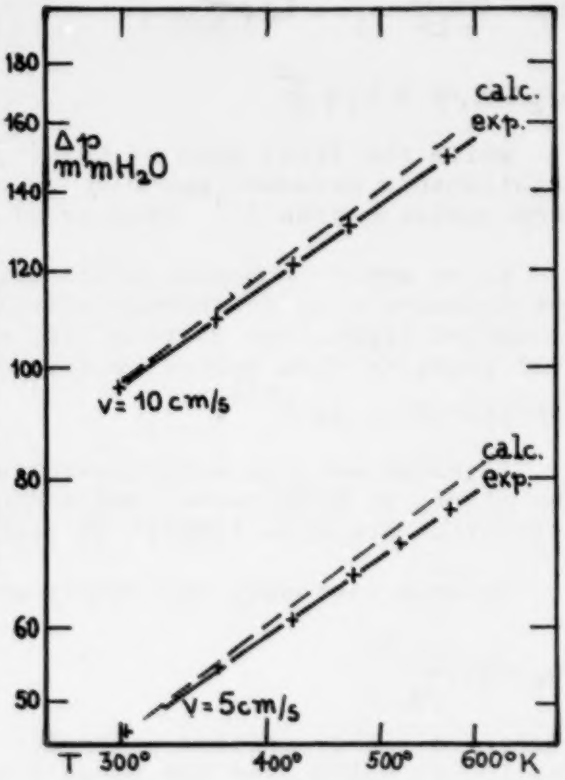


Fig. (2) - Increase of the pressure drop with temperature of a sintered glass filter of average pore diameter of 20 - 40 micron,as a function of face velocity.

Two different mechanisms are susceptible to be involved in such an effect:

a. For short tubes the pressure drop is increased by a kinetic energy term caused by the end effect involved in establishing the steady laminar flow:

$$\Delta p = \frac{128\,QL}{\pi D^4}\,\eta\left(1 + 1.2\,\rho\,\frac{Q}{8\pi L}\right) \qquad \text{or} \qquad (1)$$

$$\Delta p = k_1\eta + k_2\,\rho\,\frac{v^2}{2} \qquad (2)$$

While the first term of Eqs.(1) or (2) indicates the linear relationship between Δp and η as already mentioned, the second term yields Δp prop T^{-1} because of ρ/T = const.

b. An other mechanism which could be made responsible for the pressure drop deficiency effect is the secondary flow arising in curved pipes. From White's (18) experiments it may be computed that pressure drop caused by secondary flow alone would increase approximately as $T^{1/4}$.

Although we lack experimental evidence to decide if one or the other, or both mechanisms concurrently might be involved, an estimation could be tempted to assess their relative importance.

Gaseous viscosity may fairly well be approximated by the formula

$$\eta_t = \eta_0\left(\frac{T}{T_0}\right)^b \qquad (3)$$

where b is a constant generally comprised between 0.6 and 1.5. Eq. (3) is valid over the range from 20°C to 400°C with an error not beyond 2%. The value of b e.g. for air is 0.768. For experimental checks (e.g. as in the Figs 1 and 2) we always use the exact experimental values of gas viscosity. For the purpose of dimensional analysis alone we assumed that the viscosity of air increases with the 3/4 power of temperature.

Considering that e.g. Figs. 1 and 2 show at 600°K about 6% less pressure drop than required by the pure viscosity term alone, the solution of the following equations (4a and 4b) may yield x, the fraction of energy dissipated by one or the other of the mechanisms.

$$0.06\left(\frac{T}{T_0}\right)^{3/4} = x\left(\frac{T}{T_0}\right)^{3/4} + (1-x)\left(\frac{T}{T_0}\right)^{y} \qquad \begin{array}{ll} y = -1 & (4a) \\ y = 1/4 & (4b) \end{array}$$

The exponent y has the value of -1 for the end effect mechanism and 1/4 for the secondary flow mechanism. As for 600°K, $T/T_0 \approx 2$, from 4a and 4b one obtains about 57% in the first and 63% in the second case. Without any pretention to accuracy, one may state that both phenomena seem to play a rather important role in the pressure drop of fibrous media. They may accessorily explain, why so often the deposition on the lee side of the fibers can take place — for which laminar-flow theories offer no ready explanation.

Empirical pressure drop relationships give obviously a nearer fit to experimental data. Whitby (19) introduced an empirical function for the dependency of the drag coefficient on the Reynolds' number. Kimura and Iinoya (2o) fitted the following expression to a great number of experimental data:

$$\Delta p = \left(0.6 + \frac{4.7}{\sqrt{N'_{RE}}} + \frac{11}{N'_{RE}}\right)\frac{2\varrho v^2 L(1-\varepsilon)}{\pi g d_f} \qquad (5)$$

with $v = v_0/\varepsilon \qquad N'_{RE} = \frac{\varrho}{\eta}\frac{v_0}{\varepsilon}d_f$

The temperature dependency of the pressure drop in the Kimura-Iinoya equation is:

$$\Delta p = A + BT^{-1/8} + CT^{3/4} \qquad (6)$$

In order to fit experimental facts, an empirical term with less steep variation than $T^{3/4}$ had to be introduced. Eq. (6) fits also quite well our experiments at all temperatures and two different test gases (air and hydrogen). Fig. 3 shows the example of a carbon fiber mat tested in air.

The exponent in the Kimura-Iinoya equation allows a further estimate : all over the range $1 \leqslant T/T_0 \leqslant 2$, the function $(T/T_0)^{-1/8}$ may well be approximated by $(T/T_0)^{1/4} - \frac{1}{2}(T/T_0)^{-1}$. This could be taken as an indication that about one third of the energy dissipation by non-laminar mechanisms could be attributed to secondary flow and the remaining two thirds by the unsteadiness at capillary entrances or ramifications.

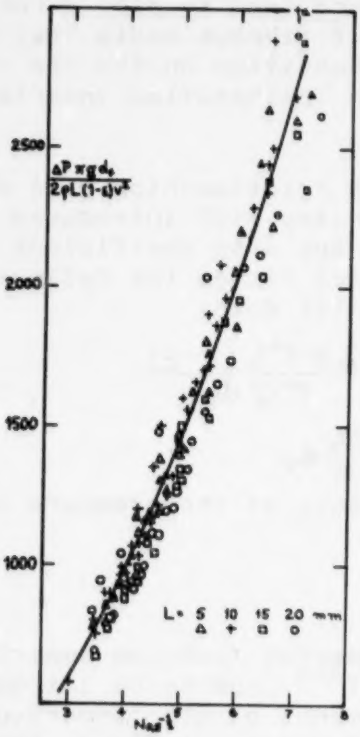


Fig. (3) - Experimental fit obtained by Eq.(5) in the case of carbon fiber mats of variable thickness, tested with air.

2. A RAPID THEORETICAL REVIEW OF THE DEPENDENCY OF FILTER EFFICIENCY ON TEMPERATURE.

This is a short brush-up of ideas already developed in the quoted basic references.

The collection efficiency E_T of a filter mat is given by the rather well known expression ("log-penetration law") :

$$E_T = 1 - \exp\left[\frac{-4(1-\varepsilon)L}{\pi d_f} \sum E\right] \tag{7}$$

where it is assumed that ΣE , the collection efficiency of a single fiber is obtained by adding up single fiber efficiencies due to separate mechanisms :

$$\Sigma E = E_d + E_r + E_i \ldots \tag{8}$$

E_d = single fiber efficiency by diffusion

E_r = single fiber efficiency by direct interception

E_i = single fiber efficiency by inertial impaction.

We do not take into consideration electrical forces because the main experiments were made with earthed semiconducting carbon fiber mats.Deposition by gravitational forces was neglected as not significant.

Single fiber efficiency by diffusion was calculated by many authors on basis of slighltly different assumptions,but all the results may be brought to the form :

$$E_d = B\ (T^{-3/2})^x \tag{9}$$

with x either -1/2 or -2/3 or -1.Anyway, E_d as a function of a positive power of T , must increase in the same sense.Following Eq.(7),the overall efficiency must grow in the same direction.

As for E_i ,Whitby's (19) approximation represents fairly well all previous experimental results :

$$E_i = \exp\left[-(\log St)^2\right] \tag{10}$$

where St, the Stokes' number or inertial parameter is defined by:

$$St = \frac{\rho_p v_0 d_p^2}{18 \eta d_f} \quad \text{prop } T^{-3/4}$$

When the temperature increases , E_i decreases, provided that $St < 1$. Thus by Eq.(7) the overall efficiency varies also in the same direction.

According to well established concepts of the majority of the authors quoted, E_r is independent or only very slightly dependent on the temperature.

The experimental and technical interest of filter tests at variable temperature follows from these basic theoretical considerations.In order to improve a filter systematically,a model must fit its real behavior.Theory allows to compute individual fiber efficiencies,but unfortunately (or fortunately) we do not have just one or two theories,but rather a dozen able to yield quite divergent numerical results.Tests at variable temperature provide a ready means to get insight in the prevailing mechanism.

To assess fiber efficiencies experimentally,the tests have to be run at constant temperature with particles of uniform size classes.This was done and published by about a score of authors. Before the automatic particle spectrum measurements were known, to gain insight in the filtration mechanism at constant temperature of a few filter mats was already a complex research project. Because of the time,expenses and care involved,it was out of question to study filtration mechanism of industrial filters this way.

As it will be shown below,tests executed at variable temperatures and not necessarily over a very wide range,represent an easy means to get insight into the filtration mechanism even with mechanically dispersed particles of non-uniform size.The tests thus become easier and more rapid.

3. FILTER EFFICIENCY AND TEMPERATURE: EXPERIMENTAL.

3.1. Particle size dependency of the single fiber efficiency.

The filter mats tested were of carbon fibers, $d_f = 8 \mu m$, $\varepsilon = 0.969$ and $L = 3.9$ mm .The test aerosol was obtained by atomizing and drying of a NaCl solution; the filtration velocity in all tests was 100 cm s^{-1} .The particle counting and sizing was done by an automatic "scintillation" (flame photometric) analyzer manufactured by Sartorius (Göttingen) and developed by Binek (21).The apparent fiber efficiencies were computed from the experimentally measured mat efficiency by Eq.(7).

The results obtained at 20°C, at 100°C (= 393°K) and at 200°C (= 493°K) are represented in the Figs.4 ,5 and 6, which also show (dashes) the theoretical single fiber efficiencies calculated by one of the most sophisticated theories : stepwise calculation of particle trajectories following Dawson (22) using the Spielman-Goren (16) flow filed for interception and inertia; the diffusion term after Stechkina and Fuchs (23) as exposed in detail by Davies (11) p. 74 ff.

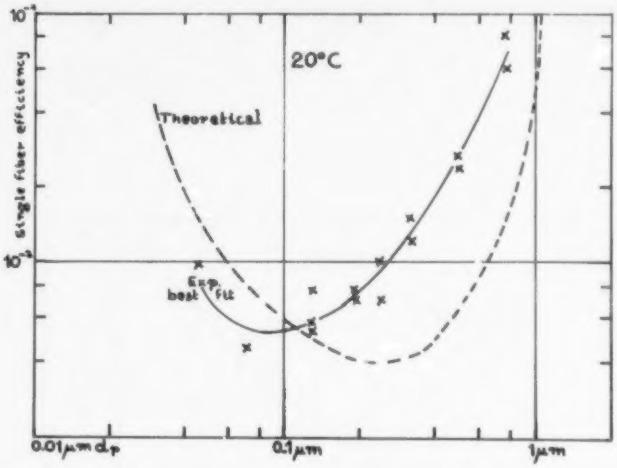


Fig. (4) - Single fiber efficiencies as function of the particle size,carbon fiber mat at 20°C;face velocity 100 cm s^{-1}.

Fig.(5)

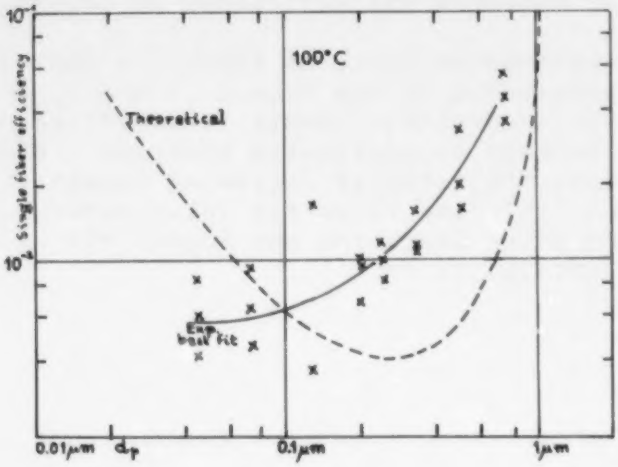
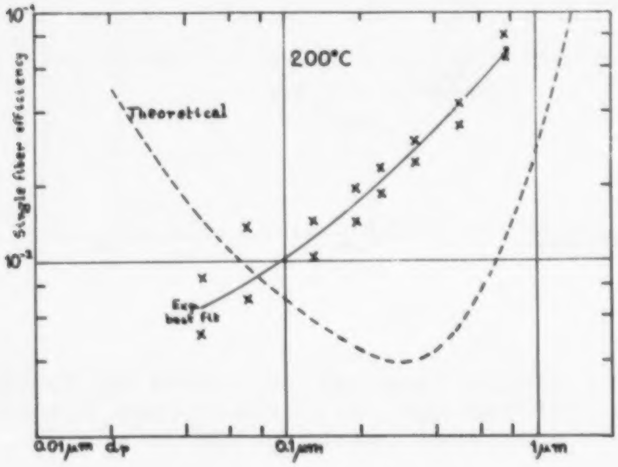


Fig.(6)



Single fiber efficiencies as a function of the particle size; carbon fiber mat, face velocity 100 cm s^{-1}.

Looking, as usual, only at the results obtained at room temperature,Fig.4,theory and experiment are in rather good fit.Both curves have the same overall trend and the fact that there is no exact coincidence should not bother us too much, as some of the constants of the theories are still empirical and depend on fiber orientation and other factors.

Up to this point,the theory seems adequately confirmed.The situation changes when looking at the results obtained at higher temperatures,Figs. 5 and 6.The minimum requested by the theory fails to show up.At first sight,we would be inclined not to trust the experiment as it often happens.But another representation of the same results (Fig. 7) does not allow this easy explanation. Up to the particle diameter of 0.13µm, one would assert that the theory is well confirmed because neither computation,nor the experiment can have any claim to very high precision.Conjecturing a somewhat different flow field and computational formula,theoretical single fiber efficiencies differing by a factor of five are often obtained (see Davies,l.c. p.64).Due to this reason,we may

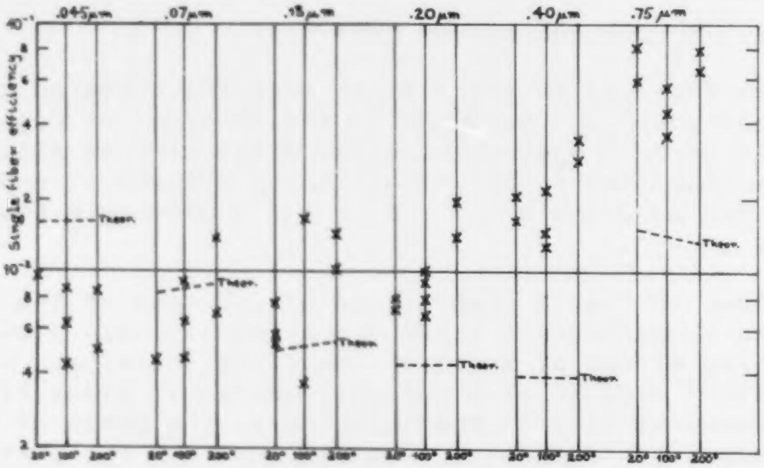


Fig.(7) - Single fiber efficiencies with particle size and temperature as parameter;carbon fiber mat,face velocity 100 cm s^{-1}.

disregard on the first sight that from 0.2 μm diameter on, the experimental fiber efficiencies are two to five times higher than the theoretical ones. More significant is the trend : at 0.2 and 0.4 μm,there appears a clear rise in the experimental efficiency, unpredicted by the theory. At 0.75 μm,the computed fiber efficiency shows already the decrease requested by the inertial mechanism,while the experimental results show a clear minimum for the intermediate temperature.

These findings, considered in conjunction with our former hypothesis that the pressure drop cannot be fully accounted for by the laminarity concept, seem to indicate that the flow field at the basis of the computations does not adequately take into account non-laminar conditions which seem to exist within the filter.

Extrapolating from this point on, we might presume that a filter in which turbulence is prevalent, will show less pressure drop and a better efficiency than a purely laminar one.Therefore, in order to develop new concepts in fibrous filtration,unsteady, turbulent or secondary-flow conditions should be sought.

3.2. Tests with polysdispersed particles.

The filters employed in these tests were the sintered glass ones already mentioned and described in the captions of Figs. 1 and 2. The test aerosol was a mechanical dispersion of silica dust, approximately log-normally distributed between 1 and 15 m with the mass-median diameter of 1.5 m and a geometric standard deviation of 4 .

The X signs in Fig. 8 show a case of decrease of the overall efficiency with temperature (slope approximately -3/4),consequently the prevailing effect of inertial impaction.The signs O pertain to a "filter" made up of a paralell bundle of glass fibers, all arranged somewhat like a brush aligned in the sense of the flow.In that way,impaction was kept at minimum and the particles could be stopped only by diffusion. As expected,the overall efficiency is rising with a slope but slightly different from 7/8.

The signs + in Fig. 8 show a case where both mechanisms seem to be involved,with a minimum of the efficiency near 400°K.

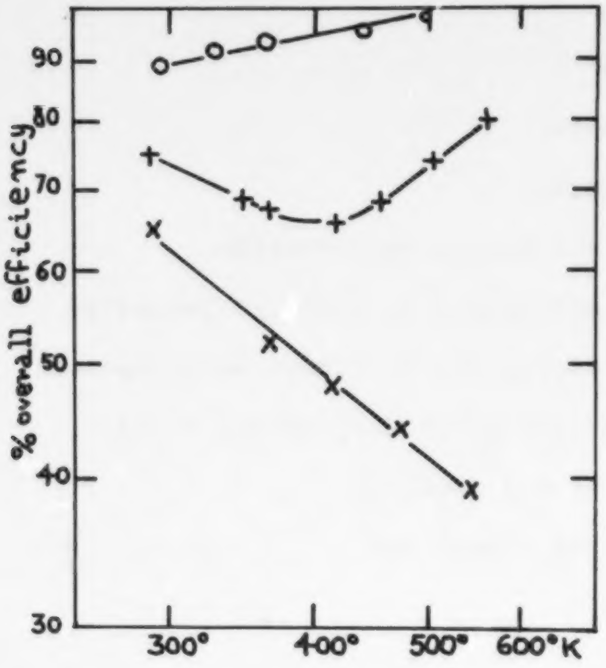


Fig. (8) - Tests with polydispersed silica dust.

LIST OF SYMBOLS.

d_f	fiber diameter
d_p	particle diameter
D	capillary diameter
E_d	single fiber efficiency by diffusion
E_i	single fiber efficiency by inertial impaction
E_r	single fiber efficiency by direct interception
E_T	overall collection efficiency of the filter
g	acceleration of the gravity
L	thickness of the filter mat
Δp	pressure drop
Q	volume flow
St	Stokes' number (see Eq.10)
T	absolute temperature
v	(flow) velocity
v_o	face velocity (velocity of the upstream flow)
ε	porosity of the filter mat
ρ	density (gaseous...)
ρ_p	density of the particle
η	dynamic gas viscosity

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WRITTEN DISCUSSION

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Dr. Benarie should be congratulated for a most stimulating talk on a very timely subject. His qualitative analysis on the temperature effects on both the pressure drop and collection efficiency should be very useful in interpreting experimental data for high temperature filtration.

My comments are primarily directed toward the following issue: Do the same physical mechanisms at low temperature for the pressure drop and collection efficiency behavior operate also at high temperature? If so, a variable property analysis is all that is necessary to predict the behaviors at high temperatures, provided that the physical mechanisms can be properly modeled at low temperatures. Professor Benarie seems to think that this is not the case,and other effects,such as capillary entrances, secondary flows and the like are operative even at moderately high temperatures. However, I would like to suggest that uncertainties in low temperature behaviors in pressure drop and collection efficiency far out-weighs the uncertainties in the variable property analysis to predict the high temperature behaviors. In other words, if we had an adequate quantitative theory for low temperature filtration through fibrous media, the variable property analysis may be quite adequate for high temperatures, with possible exception at very high temperatures.

One evidence of my speculation is the correlation of the pressure drop data shown in Fig. 3 of Dr. Benarie's paper, which, as I undertsand, covers a wide range of temperature and is based on a variable property analysis on viscosity and density based on the Kimura-Iinoya equation for the pressure drop at room temperatures.

A second evidence can be observed in Figs. 4,5 and 6 in Dr. Benarie's paper, which show the experimental fiber efficiencies over the size range at three different temperatures and one face velocity. These are the solid curves in the figures. Also shown are the best theoretical predictions at low temperatures, modified with variable properties at high temperatures, and these are the dotted curved in the figures. If we had, a good quantitative theory for single fiber collection efficiency at room temperatures, then such a theory would give up a prediction close to the solid curve instead of the dotted curve in Fig. 4. Now suppose that we use this curve as a base and extend it by a variable property analysis to 100°C, we would obtain a curve in Fig. 5 which would fit the experimental data even better than the experimental best-fit curve. Furthermore, this curve would also exhibit a minimum around 0.08μm. If the same procedure is extended to 200°C, the case of Fig. 6, the resulting curve would be much closer to the experimental data than the theoretical curve indicated, even though in this case, there is a consistent deviation from the experimental data points, which seems to indicate that additional physical mechanisms are indeed at play at this temperature level.

Based on these observations, it certainly appears that more studies, either analytical or empirical, are needed to provide a better predictive theory for low temperature filtration before any meaningful predictions can be made at elevated temperatures.

OPEN DISCUSSION

Benarie: My point was that these experiments should shed some light on the theories of the dependence of filtration on temperature. They were analyzed in three dimensions rather than two. I also pointed out that my suspicion is that some turbulence of significant duration is important. None of the actual filtration theories takes into consideration these facets of filtration. The next step should be to introduce these mechanisms, into filtration techniques in general, then we shall see a temperature dependence.

Billings: It seems to me that there is a fair amount of confusion among people in filtration about what happens in filtration. First of all, the phenomena themselves are not very well understood. They are not very well documented and not very well analyzed. The first thing desired is a phenomonological model. Before you can have a phenomonological model, though, you have to have observations. What we observe is macroscopic. We go in and measure the concentration coming into a black box. Then we measure the concentration coming out of the black box. We get back to a theory on individual fibers and individual fiber efficiency.

If I then ask one of the noted professors to explain this, they would sit down and start with a continuum theory, assume a potential flow, and imagine a streamline with a particle on it, and a fiber in the path of the particle. By doing a trajectory calculation, they can figure out where it intercepts the fiber. A lot of assumptions were made up to this point, the biggest being the matter of the particle-fiber interaction. If you look carefully at the underlying assumptions that have gotten us to this point, you will see that there are no direct measurements, by eye, of the interaction between the particle and the fiber. It all goes back to Dr. Friedlander's experiments in liquids - that is not a gas-solid case, it is a liquid-solid case. As you well know, the forces that come into play in liquids between two surfaces are orders of magnitude different from the forces that come into play in gas-solid interaction.

From our phenomonological model we can derive an analytic model. Given the analytical model, what does one do? You say that the particles will be stopped by the obstacle. Now we have a fiber efficiency and can build an analytic model. The first assumption is the agglomeration coefficient. I have given this approach to the problem many times before. It is a macroscopic basis that is dealt with. But it must be approached on a microscopic basis. The third one is a molecular basis. What you have to do is ask how effective that interaction is and what is the influence of temperature on that interaction. The same would be true of the viscosity. Instead of using the horizontal approximation, we have to look at several vertical coefficients on the equation of state at the molecular level.

Benarie: By past experiments it has been shown that filtration theories are not the crux of the problem. Perhaps all we have to do is introduce some new concepts. We ignore what was previously introduced.

Liu: Most of the theories developed are for a fiber of uniform size. If you want to translate the single fiber efficiency to a multi-fiber filter, you assume the filter to be composed of uniform fibers. In the real filters, the fibers are not uniform. So if you try to compare theory with experiment, you must try to find a particle size that will give you a good fit to the data based on the pressure drop. A different particle size may be required to fit the collection efficiency data. The question to answer is: How do you determine the correct particle size to use? That determines how much you can respect your theory.

Benarie: In this case, it was a very uniform fiber size. It was determined that it was almost one size. I would like to add that,in the theory, it has been found that a number of different fibers sizes are acceptable.

Friedlander: It is interesting to point out, even though we are using liquid systems, that we are not only having problems correlating the types of existing models in the unsteady flow or turbulent flow region, but there is also a problem at the downstream end, especially at the lower velocities. So there seems to be problems at both ends of the model.

Löffler: I am afraid I did not explain clearly enough my approach. My approach is not to consider the theory but to go into the details to analyze the fundamental phenomena.

One comment to this problem of Dr. Benarie: I can assume that in his experiments, the difference between experimental and theoretical values should be an effect of electrostatic forces. This is because you used sodium chloride particles produced by atomizing. All these particles have charges. The tendency of this electrostatic effect of increasing 10 or 20% is minimal. This would mean that increasing the viscosity would create an effective decrease.

Benarie: Therefore we do first the electro-conducting filter. Your explanation does not correlate with the pressure drop. I consider the pressure drop easier to measure and more indicative of turbulence than the deficiencies that are being criticized,because electrostatic effects, adhesion and impurities can be changed by temperature. I need not grade my arguments,but if I woud, I would give myself a 9 out of 10 for keeping pressure drop and coefficients.

Oder: I think Dr. Benarie gets a 10 for candor.

Cooper: One thing that strikes me is that the experimental results are substantially higher than the theoretical, above a few tenths of a micron. Which would probably suggest that we are not having accommodation or rebound on the fiber. One of the things I wanted to know was what would be the Reynolds numbers of the fibers in the flow. Could we expect a wake?

Benarie: Do not be surprised if the results are 2 or 5 times different from the theory. Because if you take the models ,97 of them, and compared the different flow fields within the same theories, discrepancies among theories are present. So a person chooses the theories they desire to make the best fit.

The Reynolds number will be discussed in the panel discussion. If you have a tube, what is the Reynolds number; the thickness of the wall or the tube diameter? Why is the fiber Reynolds number so important? There is not a great difference. With a porosity of 90% there is a factor of about 2 we have to deal with. The right Reynolds number is around unity. Alone, the Reynolds number can not explain the turbulent development. But we can cover that in the panel discussion.

Billings: My point is that you are assuming that the Sutherland approximation to a Kuwabara approximation to the flow field is not going to lead to a set of problems. I think we need to look at the accommodation coefficient for the molecular material on the solid. It is at the molecular level that the effects come in. The viscosity arises because of the accommodation coefficient of the solid for the gas. That is a non-stick assumption on the solid which is probably not true. I think you have to look for temperature dependence on the equation of state to look for the change in the viscosity.

Benarie: We analyzed the viscosity effect and about six different possible effects. The non-slip coefficients explain only about 1% of the observed effects.

PANEL DISCUSSION

Subject: "New Directions in Particulate-Gas Separation Research"

Moderator: Benjamin Y. H. Liu

Panel Members: Teoman Ariman, Michael M. Benarie, Dennis C. Drehmel, Robin R. Oder and David S. Scott.

B. Liu: First of all I would like to congratulate Dr. Ariman for the excellent job in the organization and guidance of this workshop. The purpose of the panel discussion is to look at the new directions in particulate-gas separation research. I think the panel is well represented with experts in the various research areas. It will be a ten minute presentation by each member of the panel. Then a general discussion with the participation of everyone will be held.

I would like to begin the discussion by giving a few of my general comments, informal questions, and suggestions and finally some of my opinions.

First of all with regard to magnetic separation, I really don't know much about magnetic separation but one thing that has bothered me is that no one in this conference has dealt with the problem of energy needed in order to grind a particle down to the size to separate out the mineral impurities and the sulfur. So I would very much like to see Dr. Oder address the question of the energy that you have to put in the system to grind a particle to two or three microns in order to separate out impurities. With regard to electrostatic separation I would like to say that, if my memory serves me right, a few years ago electrostatics was the whipping boy for the blame of the experimentalist whenever he could not make an experiment agree with the theory in aerosol studies. Usually the first thing you point out is electrostatics. So I am very happy to see that now at least we are putting electrostatics to good use and I think the electrostatic is no longer a black magic as it once was. It has begun really to become a science. People are beginning to understand and study it and know methods of measuring the charge and neutralize the charge and handle the electrostatic problem much better than they did before. So,I am very happy to see that advances have been made in the electrostatic area.

Now for the case of the acoustic interaction I think Dr. Shaw had pointed out clearly that the problem seems to be that we don't have a good handle on the theoretical coagulation coefficient for particles in the acoustic field. We know that it is very expensive to perform full-scale testing of these ideas. Therefore, sound theoretical understanding of the basic phenomena involved is very, very important.

So at this point I would like to put out a pitch to EPA and NSF for the importance of the fundamental research. I think the reason that we have to devote attention to the basic research is that by understanding fundamental phenomena at a microscopic level we can avoid the very expensive and costly mistake that we could make by using a larger scale apparatus. So I am very glad to see that the NSF and EPA are joining hands in sponsoring this workshop. Obviously EPA has its own mission to meet so therefore they cannot sponsor too much of the fundamental research whereas NSF certainly is in the business of supporting basic research. I would like to see in the future that more fundamental university research programs be supported because it is very important in the long run. Furthermore the fundamental study may turn out to be also the most economical way of solving some of these problems.

So these are some of my very brief comments. At this point I would like to turn to Dr. Benarie.

Benarie: I begin this discussion with an answer to a question. You see quite clearly that my purpose of presenting some research in three-dimensional channel filtration was to rock the boat a little bit. I continue rocking the boat with the following.

One of the most important theories of pressure drop is the channel theory. We do not have a channel theory of efficiency; we speak only about fibers. I would like to draw attention to the fact that there are holes between the fibers. Because the air passes through the holes, the particles are captured by the fibers. I find that by drawing just the holes in front of the channels, and, instead of looking at the fibers, by looking at the channels of air coming in, you immediately have turbulent agglomeration. We have no where taken into account the turbulent agglomeration within the filter.

Next, we have here the effect of secondary flow, because it disturbs us, we don't take it into account. Thirdly, at all beginnings and ends of channels we have an end effect which is turbulent-like; this also is not taken into account.

I suggest that what we need theoretically is a channel theory of sections. My second suggestion is that we try filters with different shaped fibers to try to enhance in-filter turbulence and take a step forward in meaningful turbulent filtration by mass. Currently we do most filtration through fibrous media which has not straight but curved sides. More advanced technology would allow us to make any micro-fiber shape. A turbulent pressure drop is less energy-wasting than a laminar pressure drop. Perhaps we can get more efficient and more energy conservative fiber shapes. A number of fiber capsules with rough and smooth fibers are available to us for full testing. So exterior fiber geometry still has the potential to bring a renaissance to this old art of fibrous filtration.

Drehmel: I guess like Dr. Billings I have to start with the microscopic level and work backward. We of course at EPA are mission oriented. We have to provide demonstrated technology to control particulate emission from stationary sources. And unlike some of the other goals that we might pursue such as understanding the theory better we quite often get ourselves into the position of knowing a piece of hardware works and then asking ourselves, what happened.

In order for us to make some sense of what is happening with the equipment and also to understand how it will fit into the total picture of particulate control equipment,we do get involved with theory. But our goal is to provide technology. So let us spend my few minutes looking at where problems are, and some of the solutions we have taken so far.

So far we have three types of technology that are available: filtration, scrubbing, and electrostatic precipitators. We know that under certain circumstances at least an efficiency of 99.9% exits. We have heard people mention today that filtration efficiency was 99.99%. Collecting the particles isn't necessarily the problem. It is the circumstances under which you have to catch these particles that give us some difficulty. The precipitators have problems with resistivities that are too high or that are too low, the pressure drop in scrubbers is too high and the filters are too big. And so research to this point has taken the attack one of trying to improve standard technology and also to go out and find new types of technology that could be brought to bear. The types of research we have looked at are pre-charge chambers or 2 stage precipitators in order to get around resistivity problems. We looked at conditioning with agents such as SO_3 or a sodium carbonate. We have looked at baghouse problems where we have size limitation. We are looking at high velocity types of filtration where we go to air cloth ratios of 3 to 10. We like to look at an air cloth ratio of 1000. We also would like to look at media that were not limited to less hostile environments. We would like to look at media that could withstand high temperatures, we would like to go to 800°C, not just a couple hundred degrees. We would also like to be able to deal with environments that have acid gases, sulpheric acid, HF, we have found a great deal of work on that.

Since the point has been brought up about filter geometry, we have funded research looking at smooth fibers vs. trilobal fibers. And we found in fact that the trilobal fibers worked much better. For scrubbers to get around the pressure drop problem, we have looked at the device called the flux force condensation scrubber. We bring to bear condensation on the particulate which helps remove particulate both by growing it and also by bringing to bear diffusional and thermal mechanisms.

So this is where some of our research has been oriented on conventional devices and trying to take advantage of new mechanisms. We have funded research in several areas, we have looked at sonics and spent some money on magnetic, but primarily we have looked at the combination of electrostatic and other forces. And for example we had funded some research at MIT to look at systems with charged droplets and particulates. That is where the time constant theories of Dr. Melcher came from. We asked Doug Cooper when he was still at GCA to look at evaluating electrostatically augmenting devices. And you have also heard about these other projects with Battelle Northwest and TRW. We have also talked to George Fielding to do some work with us. We are currently funding Mike Pilat to look at his charged droplet scrubber. We felt that electrostatics provides a very effective force in the control technology of removing fine particles.

As to where we have research needs in the future, rather than talking

about some of the theoretical research needs (I am sure it will be brought up more by the other members of the panel) I would like to take a different attack and say that primarily we have talked about particulate-gas separation from a conventional point of view. We have talked about them in terms of a stationary source with temperatures or pressures that we are used to looking at. We find that the suspended particulate, where we have problems in not meeting particulate standards, is primarily due to sources that are not enclosed sources. They are what we would call either fugitive emissions or fugitive dust emissions. Fugitive emissions would be the type of thing where a source is not completely contained, not properly hooded. Because of the process we have had some difficulty with an open electric hearth, for instance, where we have some emission around the site or sintering machines which are not completely enclosed. We would also have fugitive dust emissions where dust has been deposited from an industrial process which is now a roadway in a plant or it is part of an open mine face, or it is where work is being done in an ore deposit. Even more troublesome are the types of things where we have asbestos from brake linings or we have particulate from the use of tires. We find that all these things contribute to this suspended particulate and we find a
this area. We need to have research oriented to the control of particulate from non-enclosed sources. Also we just heard that we are going to use more coal. One of the ways we go about using more coal is to gasify it or burn in a fluidized bed combustor. In either case you can make these processes much more efficient if you can use a gas turbine. A gas turbine operates at high pressure and it would be necessary to remove the particulate in advance to the gas turbine. We are talking about particulate control at 1500°F (800°C) and 10 atmosphere pressure. Again we suddenly find that there is no technology available. We have various types of cyclones that can be used or have been talked about. They do not provide control technology below 1 micron. In order to provide particulate control for both the turbine and environmental protection we will need research in this area. So I would see that much of the applied research for the near future will be in line with conventional devices. We will work with conventional devices to make them solve the problems but innovative research has a large application to areas where we have no technology; to the high temperature, high pressure control problems and to the fugitive dust emission area.

Oder: Coming from an engineering firm, I came prepared to address many of the topics already treated by Dennis. So now that he has given his presentation, I am free to come back to magnetic separation and other areas.

Before doing that, however, I would encourage the funding agencies to fund scientists and engineers so as to bring these people into the heart of the problem together. The scientist, from the engineer perspective, is apt to be paranoid and to think that the engineer is limited in his vision and is a negative influence. The scientist is concerned that the program may be stopped at an early stage because of peripheral constraint that the scientist is not even interested in. I think that this viewpoint is wrong. I also think that overconcern with engineering economics can be a negative influence. So I would like to see a program that studies feasibility from a creative perspective at a very early stage in any developmental program and magnetic separation is one area that certainly could use that kind of insight.

Now let me turn to magnetic separation. Gas-particulate separation in this area is new. Basically, the only experimental evidence has been gathered by Drehmel and Gooding. So other than their work, there is conjecture and a great deal of that. The technology of HGMS that they are studying has been successfully developed commerically for wet mineral beneficiation. However, in the area of dry separations it is in its infancy.

Now I want to also remind you that the magnetic technology is severely limited in its applicability. It doesn't have broad general applicability. It is usable only in systems which can be magnetized. The stronger the magnetic material the more suited it is to magnetic separation. Now with that idea in mind I would also suggest to you that when looking for applications of the technology in gas-particulate separations you must address the problems associated with high velocity processing. In order to be able to afford HGMS technology high material throughputs are required. So whether it be in areas of removing mineral matter from coal or whether it be removal of suspended particulate, such as in the case for the steel making furnace off gases, you need to have large capacity systems to even think of the technology in terms of air-particulate separation.

There is a plus going for magnetic separaston which hasn't been mentioned here at all. In those areas where the materials are strongly magnetic and where magnetism is iron based, the technology can be employed, in principle, up to 1400°F, which is the cure temperature of iron. So in principle you are talking about potential applications up to that temperature. So if one has applications where magnetic forces will work, he is not limited to low temperature applications only.

Now with regard to the question about the cost for pulverizing; first of all, I did not want the emphasis of my talk to be on coal cleaning. Rather I wanted coal cleaning to serve as an example of an area where one could potentially make strongly magnetic materials so that an application of magnetic methods of gas-particle separation similar to the steel furnace application could be studied. But in that particular context, the costs breakdown the following way: pulverizing including crushing is about \$2 or \$2.40 a ton based upon a conservative design. Coal separation might be done for less. The capital cost for the desulfurization circuit considered here including all reactors and including the magnetic separator, is about \$105 per ton per hour of clean coal product. Operating costs are estimated at \$7 per annual ton of clean coal for desulfurization only. And if one wants to store the clean coal, then the cost for compacting the pulverizing coal is about \$4 per ton. So, the costs for pulverizing are expensive, but I don't think that is going to be a major consideration in limiting the application to dry processing of 200 mesh coal.

Now let me say something about fibers. The magnetic separation technology has perhaps been studied more extensively from the standpoint of choice of collection material than has the electric filtration and that surprises me. It is known that the shape of fiber, the local surface curvature and the deployment of the fiber in the packed bed is crucial and extremely important to magnetic separation. Kaolin processor have learned how to selectively remove particles on the basis of their size. The bulk strand diameter is important in collecting large particles over 8 microns in size and fine size surface imperfections on the strand surface are important to the removal of materials in the 0.25 micron size range. This can be confirmed by etching the imperfections away.

I will now list areas in gas-particulate separation that I think are very interesting.

I would like to see more information about the strength of the coagula that are formed utilizing the agglomeration technologies. The idea is that electric or magnetic separations might be aided with use of appropriate seed material. For example, nonmagnetic particulates could be seeded with magnetic particulates to form a coagulum to be removed magnetically. What is the strength of that coagulum? What strength turbulent forces would break it up? With regard to fluidized beds I feel there is something very fundamental that hasn't really been touched on and that is the stability of fluidized beds. I do feel naively, but cannot prove to you, that it is possible to introduce electric and magnetic forces to promote stability in fluidized bed processing. By stability I mean no bubbles, no mixing in the bed whatsoever. Particles which are absolutely fixed in space would be useful as filter elements.

I am curious to know if anyone has tried the use of magnetic forces to suppress arcing in electrostatic precipitators? One thinks of capacitor suppression of electric arcing in an automobile ignition system and of magnetic methods for suppression of arcs in submarine battery systems. These devices are magnetic in nature and they blow the unwanted arc away. Just a curious point, there might be some magnetic way to suppress that arc in electrostatic precipitators.

The last comment that catches my imagination is the use of foaming technology to capture fly ash from coal combustion products in flue gases. I wouldn't suggest the same foaming agents that have been discussed here. But if you could precede the precipitators with some kind of a foaming agent that was liquid at the operating temperature (a few hundred degrees Fahrenheit) and you could blow large bubbles (a quarter of an inch or so in size) then you could encapsulate many millions of micron size particles per bubble. Then by capturing and destroying those bubbles, you would have a very nice way of capturing particulates which are otherwise very difficult to control. That is my list of suggestions.

Scott: I obviously would be glad to talk about acoustic dust conditioning but I think I will not. Rather I will title my comments something like "Notes of a Workshop Watcher". And I want to discuss two themes under the "workshop watching idea".

First is the theme of technology transfer. I think at this meeting we have a set of people who are working to better understand the fundamental mechanisms of those processes with which we deal in particulate-gas separation. Needless to say, to do that right they are isolating various sub-mechanisms and so, necessarily they can examine only a few. It is, therefore, very difficult to thoroughly determine the macro-mechanism of any given process. But it is clear that this fundamental micro-mechanism research is the touchstone of all the work we are doing. And I agree entirely with Professor Liu that this must be encouraged because it is the only way we can really achieve breakthroughs in the second level.

The second level is when people perform large scale evaluations, or overall evaluations, of total processes. Things like power requirements, resident times, capital costs, collector efficiencies, etc. are examined. I guess the comment I would make here is that these have to be done as completely and honestly as possible, using similar criteria in each research group - particularly on collection efficiency. I think there is obviously an area of research that we haven't talked about very much here, and that is the development of techniques for particle sampling, in particular, very fine particle sampling. And maybe I could suggest that Grady Nichols or Michael Pilat (who at least in times past, I think, have worked on fine particle sizing) might make a comment later on.

The next step is to commercial operation. This is where technology transfer becomes more difficult. Let us presume that we have a process which has been developed and works well. Somehow it has to be used. And the fact is, that the best way for it to be used is by our free enterprise system; that is, you have to have manufacturers which are picking these things up and carrying them on. But then we are into proprietary interests, practical marketing constraints, etc., which sometimes are not at all scientific but are nonetheless very, very real. The technology transfer from fundamental research to second level, or more applied research is a process which works. It works from conferences like this. But once you get the process into equipment manufacturing - with sales, maintenance, and many other forces at play - then I am not very sure how it works. That is why I introduce the question. I am really suggesting that in the next hour, there might be some comments on technology transfer barriers. Maybe it is not as serious as I think it is. But I think the difficulty of commercializing the university (as distinct from industrial) scientists' ideas is very real.

My second theme under this "Notes of a Workshop Watcher" discussion, is a theme on the interrelation of processes for accomplishing gas-particulate separation. This comes from two observations of the last couple of days. The first observation is many approaches and perhaps the most promising approaches - how one judges the most promising approaches is difficult - are hybrid, or they involve physical phenomena of several different types. My own field is acoustics which, of course, is a conditioning process and thus demands that there be some other collector. The point is that there are many systems. I think it is unlikely that any single process as it was defined at this conference will find its way to wide industrial application - unpollinated by other ideas and processes.

A further observation related to the interrelation of processes is that new particulate-gas separation applications are popping up like mushrooms after a warm spring rain. I think Dr. Drehmel has mentioned this, and others have seen it. But it is something that I have certainly been struck by during this workshop. And these new applications are coming with obviously a set of new requirements. I think a lot of processes are hybrid and will continue to be hybrid. Moreover, applications emerging which will yield requirements we cannot anticipate. All leads to a comment about the credibility of a process. I think that none of the processes we have been discussing are yet credible. I think, in fact, that we don't even know how to judge them so that they could be defined as credible. For example, we don't even know yet what the dust load requirements at the inlet of a gas turbine are! We don't know yet what the cotton processing requirements will be. So how can we say that processes designed to satisfy these as yet undefined requirements are credible or not? I think the question is not the credibility of the process, rather it is the credibility of the study.

Finally, I come back and wrap all this up. First of all, we need fundamental work which is credible. Secondly, we need results - especially those that analyze the macro-processes to be as complete as possible. In that sense it is as important to identify the faults of the process as it is to identify the good parts. Perceived negatives may turn out to be strengths after the applications are clarified. Thirdly, we have to identify all the criteria. I have mentioned several: power, dust load, size distribution. If Jim Melcher is going to be at the next conference you know for sure you have to evaluate residence times for him to give you his ear.

My last comment with regard to all of this is that the pride in one's work is obviously essentialy for good work. I suppose it is only natural that those who are working on a particular process have more faith in that process than do their peers working on other processes. We are all at fault, sometimes, in this. We've got to be careful that this natural pride does not change into something that sounds more like shrill advocacy. I think intense,well - articulated discussion on the interpretation of the physics of processes is very productive. Yet I think we must avoid a posture of shrill salesmanship for specific processes. As such, it could be very counterproductive for all of us.

Ariman: I think I would like to concentrate on mostly electrostatic filtration in this short presentation. I am sure you noticed that there are several interesting things happening and many more difficult problems remain to be solved in electrostatic filtration. When you talk about electrostatic filtration you have to differentiate two areas. Namely electrostatic fibrous filtration and electrostatic fabric filtration. If you noticed in this workshop and literature we really are concentrating on a mathematical modeling of electrostatic filtration by fibrous filters. Professor Löffler, Dr. Zebel and number of other investigators have been looking into the problem of a single fiber and a single particle and are trying to make a mathematical model for it. Hopefully maybe for specially built fibrous filters this model could be used. However, when you look at the fibrous filters you use in your air conditioner, furnace, etc., we are really facing a much more complicated situation than a model with a single fiber and a single particle. On one hand, we have a very basic and also quite mathematical single-fiber model. On the other hand, we have almost thing but a black box as Dr. Billings indicated.

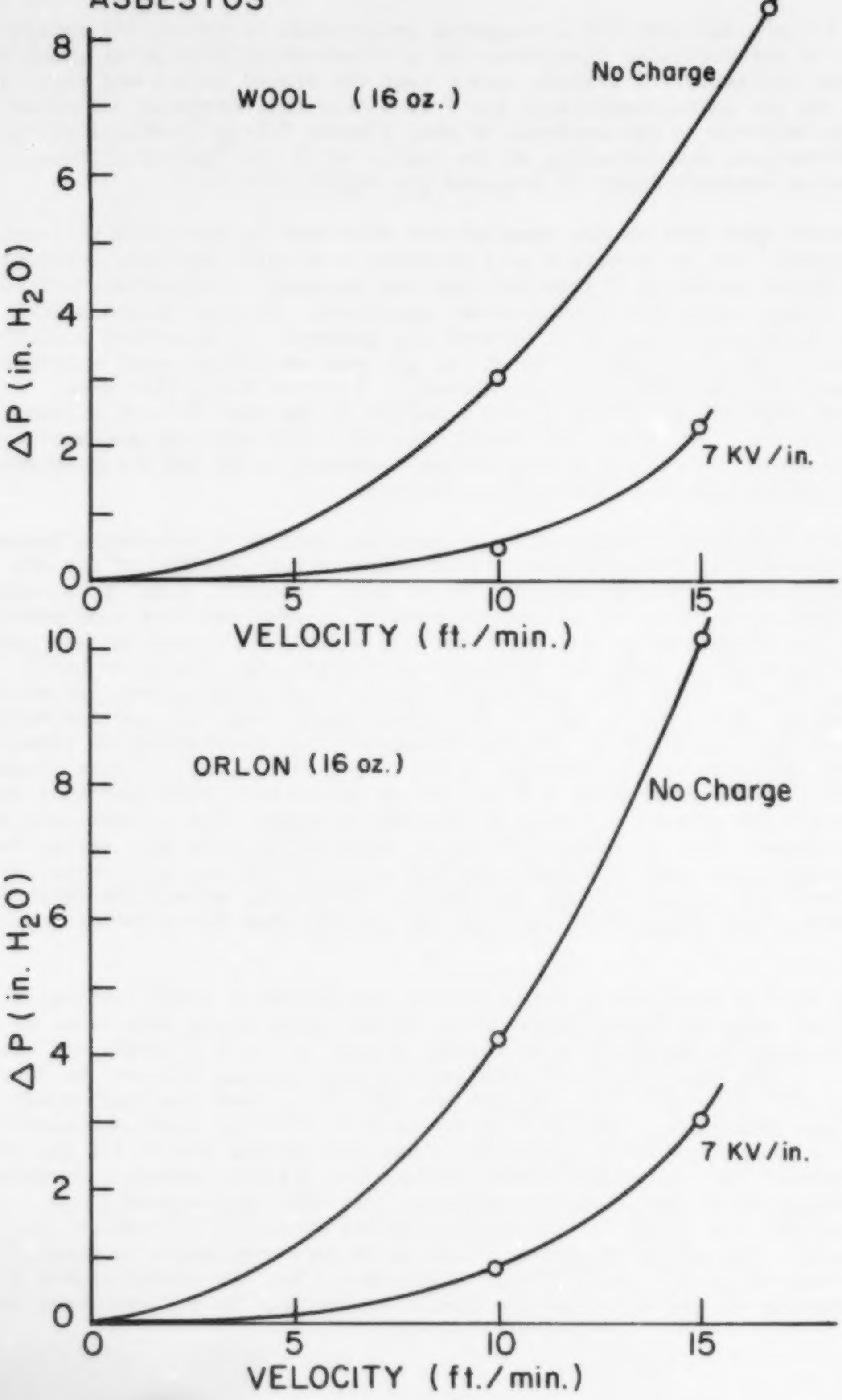
Furthermore, we have been doing a lot of testing in particulate collection with fabric filters. Then, by mostly using the experimental data we have been developing empirical or semi-empirical formulations for the collection efficiency and pressure drop in fabric filtration. It seems to me the time has come, particularly when an external electric field does exist, to take a systematic approach to the very difficult problems of fibrous filtration, in the following way (1) a single fiber model (2) a group fibers in the same row (3) fibers in a number in parallel rows (4) a staggered array model of fibers (5) consideration of fibers in perpendicular directions (6) a three-dimensional model which would be the most realistic. I readily accept that the stated models and particularly the last two are quite complicated but I feel that this proposed way could be a reasonable approach to the analysis of real fibrous filters. Hopefully, with some modifications and depending on the series of it for fibrous filters, a corresponding approach could be utilized for fabric filters.

I would also like to see experimental work hand in hand with the analytical investigations. It is certainly very attractice to carry out pure theoretical work in fibrous or fabric filtration with fourth-order, sixth-order, ordinary, partial, linear, non-linear differential equations. We also have powerful computers which give us a lot of interesting numbers. We also have a capable draftsman to make nice figures; so we can get some excellent publications out of all these in reputable journals. However, I do not think this pure theoretical approach with very little relation to the real life of filtration is a healthy one. Therefore, we should make sure that whatever mathematical model we propose, we provide a very valuable support to it and verification of the model with a suitable experimental program.

On the other hand yesterday, I am sure you noticed a remarkable decrease in the pressure drop in the electrostatic filtration by APITRON of the API and our bench scale test equipment at Notre Dame. However, some of my colleagues tell me "well, look, you are trying to convince us that you save some energy by reducing the pressure drop, but how about the additional energy you are putting into the system by charging the particles or charging the filter or both? What happens to that?" Well, if I may have a couple of transparencies, let me show you something. Based on our bench scale experimental results, and the full scale field tests by Dr. Helfritch in industrial dust filtration, we obtain a substantial decrease in the pressure drop. First of all, why is this happening? We studied the dust cake under the microscope and we have found out that the dust cake differs quite a bit from the noncharged cake. The charged cake looks much more porous than the noncharged case. When you do have more porous dust cake, you will have less resistance against the flow and you don't need too much power to pull air or gas through the system. Hopefully, we will be able to produce some three dimensional pictures and publish them for a better understanding.

Now if I go back to a power situation for the bench scale testing, we have up to 80% decrease in the pressure drop. In our bench scale experiment we have usually a corona formation by a DC charger around 5.5 or 6 kilovolt per inch (kv/in.). In our experiments we obtained the best results between the 7 or 8 kv/in. In the graph of power, you can see that it is less than half a watt. Now I am not speculating that will be the case in the full scale application, but at least our bench scale experiment shows that we use almost all the power for the exhaust fan system and almost nothing for electric energy. Therefore energy savings up to 80% could be realized. In 1976 approximately one billion dollars was spent for the power required in fabric filtration in this country. Therefore, it appears that it is very worthwhile to conduct further research in the electrostatic filtration. Now the second aspect I would like to discuss is the effect of the fabric structure. We all know when energy

ASBESTOS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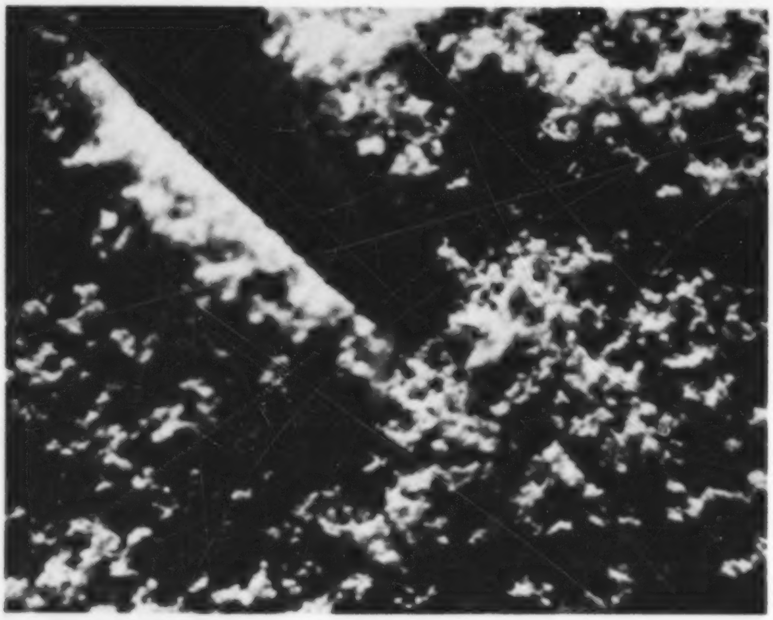
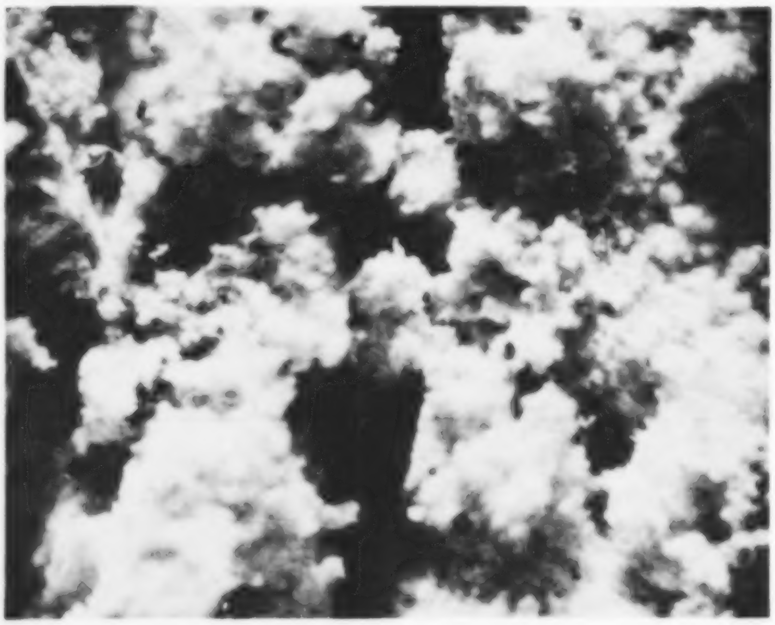


UNCHARGED ASBESTOS CEMENT DUST
DEPOSITED ON FABRIC - 5X MAGNIFICATION



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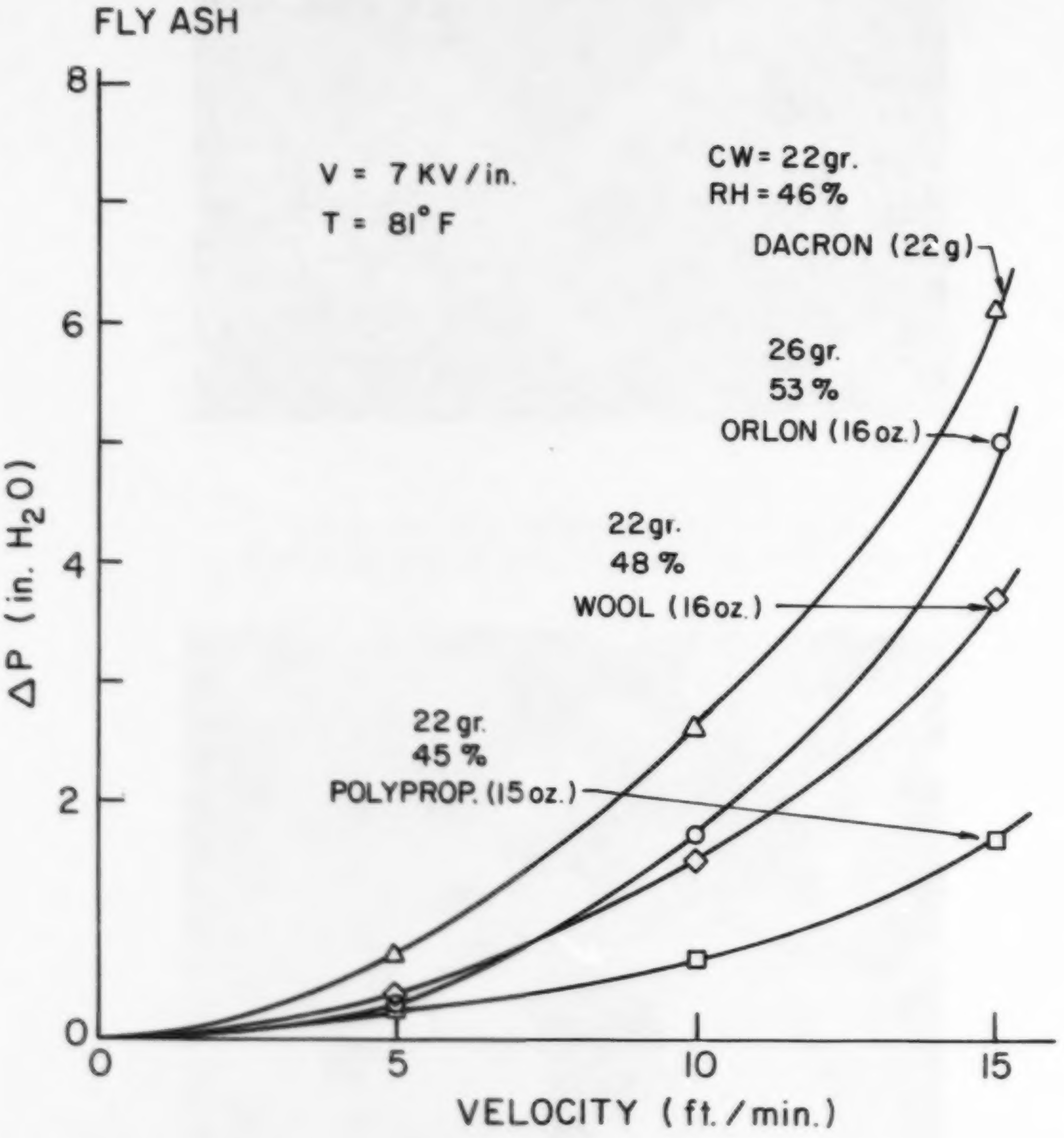

FLY ASH
V = 7 KV / in.
T = 81° F
CW = 22 gr.
RH = 46 %
DACRON (22 g)
26 gr.
53 %
ORLON (16 oz.)
22 gr.
48 %
WOOL (16 oz.)
22 gr.
45 %
POLYPROP. (15 oz.)
ΔP (in. H_2O)
0
2
4
6
8
0
5
10
15
VELOCITY (ft. / min.)

was very cheap nobody cared about possible savings through a lower pressure. The second transparency shows the effect of fabric type for coal fly ash dust samples and you can see that there is a considerable effect of fabric type on pressure drop. So I think experimentally we should look into the fabric structure in filtration with or without electrostatics. I should point out that Dr. Drehmel and his associates were the first to study the effect of fabric structure in the filtration process.

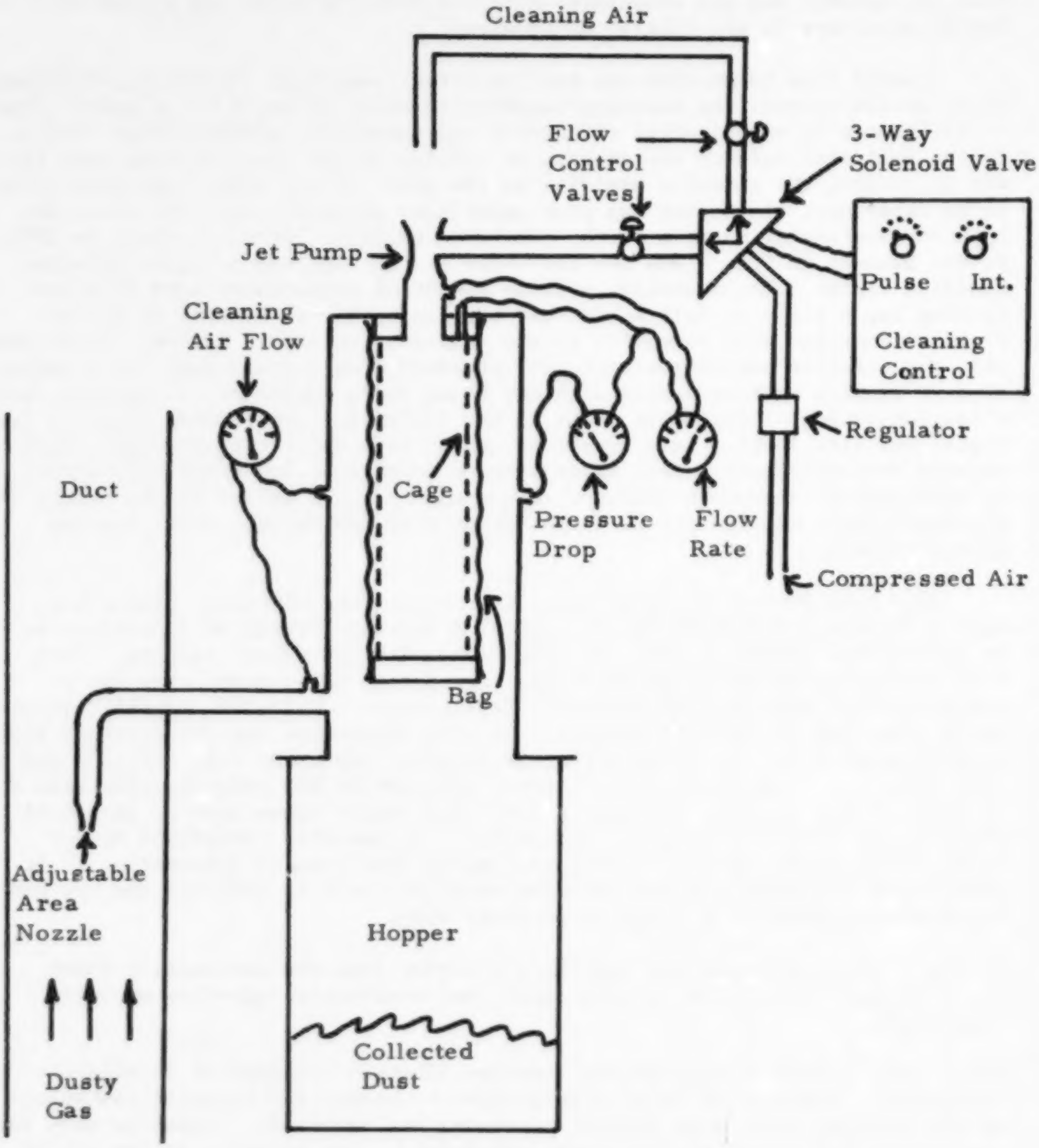
I would like to mention one more important aspect of filtration, humidity. If we do not control the relative humidity it seems to me there is quite a bit of difference in experimental efficiency and especially pressure drop results. In the full size outdoor experiments or testing in the field we have very little way to control the relative humidity in the air. It may rain a day just prior to an experiment or you may run your experiment after the rain and there may be some substantial changes in the relative humidity. Until my visit in 1973 to Dr. Drehmel in EPA, I was not too aware of the importance of the relative humidity factor. Incidentally, a large number of researchers have been conducting bench scale or full scale test results in the laboratory or in the field without too much attention to the relative humidity. However, it is seen that the relative humidity affects the pressure drop a great deal for a number of dust samples we have tested with our bench scale equipment. It appears that a reaglomeration of particles prior to the filtration phase takes place at the higher humidity level. As a result it can be seen under the microscope that we usually have more porous dust cakes than we do have at lower humidity levels. As indicated by Professor Löffler, this reaglomeration is due to the change in adhesive forces between the particles themselves and between particles and filter surface.

This time instead of bench scale testing we used miniature single bag dust collector designed by Dr. Helfritch of American Precision Industries as in the figure. There is just one single bag with jet pulse cleaning. With this single bag equipment we were able to measure the pressure drop for a number of dust samples with different fabric bags at different humidity levels. Let me show you a couple of results. In this figure you see the pressure drop versus velocity for the three different humidity levels at 24%, 46%, 84% for furnace fume dust. Again you see a linear decrease in the pressure drop with an increase in the relative humidity. Now, this really shows that we should be aware of the fact that if we do not control the humidity conditions in our tests, it might be quite different to compare test results. Secondly, it is possible to condition the gas by water spray not only to cool off the hot gas but also to provide a decrease in pressure drop.

B. Liu: Before we begin the general discussion from the audience, I think Dr. Robin Oder would like to make additional statements regarding magnetic separation.

Oder: Yes, I want to address the question of power consumption in magnetic separation. There seems to be a reluctance to accept the magnetic technology on the pretense that large amounts of energy are required. I want to make the suggestion that you make that comparison fairly. For example, in the use of electrostatic precipitators, one might estimate how much energy is consumed

FIGURE
Miniature Single Bag Pilot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Flow Control Valves
3-Way Solenoid Valve
Jet Pump
Pulse
Int.
Cleaning Control
Cleaning Air Flow
Regulator
Duct
Cage
Pressure Drop
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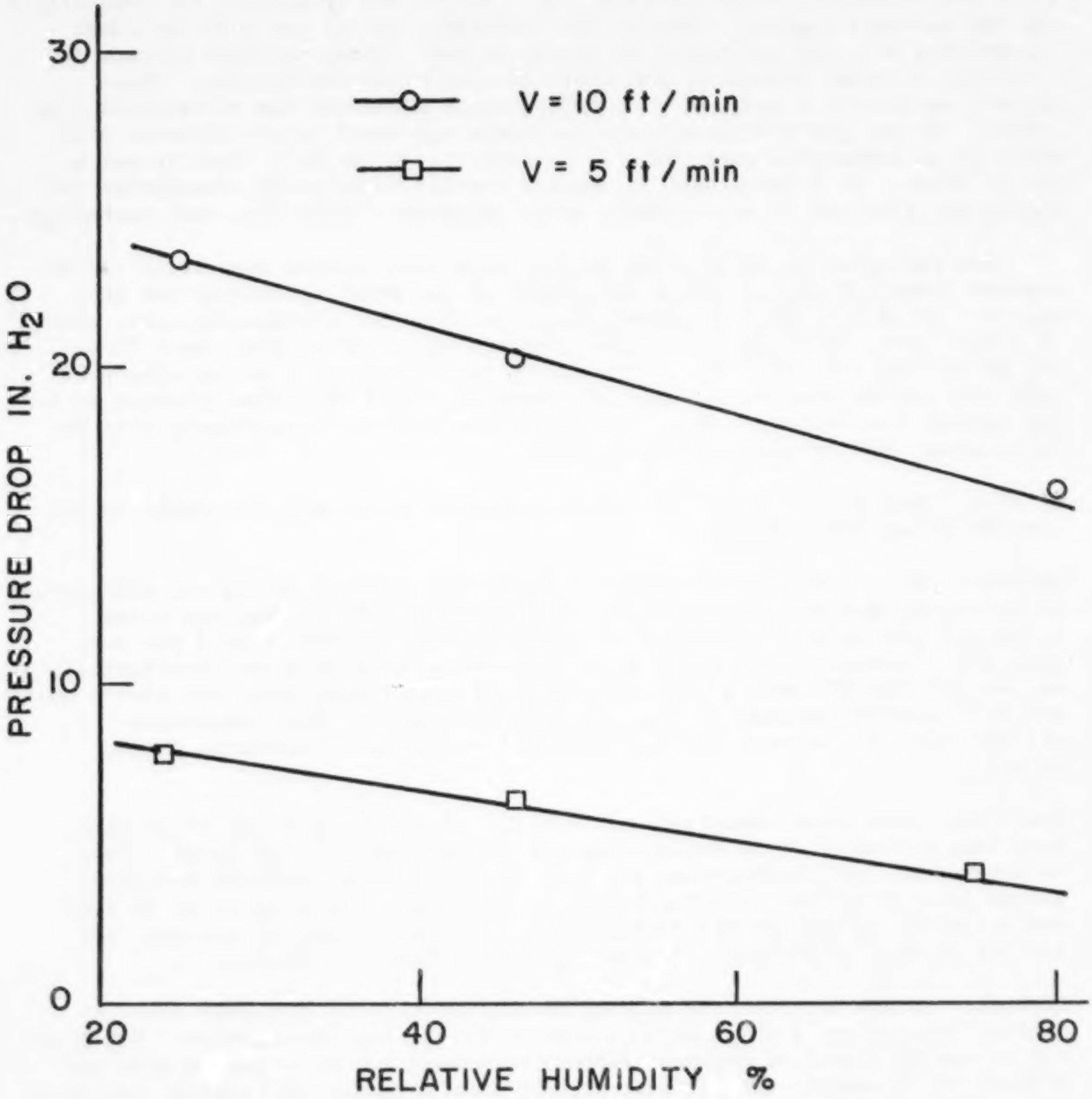
Hopper
Collected Dust
Dusty Gas


V = 10 ft / min
V = 5 ft / min
PRESSURE DROP IN. H_2O
0
10
20
30
20
40
60
80
RELATIVE HUMIDITY %

in terms of kilowatt hours per ton of material that you capture.

Now, the only commercial technology in high gradient magnetic separation is in the processing of kaolin clay. As I showed you Wednesday, the separations are 500 kilowatt magnets. That sounds excessive, but if one puts the power consumption on a per ton basis, it is not so bad. Those machines operate in a variety of modes depending upon what the clay processor desires. These magnets operate in a range of 9 to 100 kilowatt hours per ton of material processed. If you put that in dollars and cents and use 2.4¢ per kilowatt hour which is an industrial rate, that is 21 cents to $2 per ton. That is not a lot of money. So I think that if you are worried about power consumption for magnetics, then put these estimates on a comparative basis with your technology.

Now the point to pulverizing is that liberated mineral impurities can be removed from coal with magnetic separators or any other technology for that matter. It is felt that by pulverizing from 70 to 80% through 200 mesh, which is called power plant grind, one can remove mineral sulfur from about 80 to 90% of our eastern coal. So extraordinary pulverizing to 1 micron size probably will not be required in physical cleaning. Such extensive grinding is for the future, for selected coals. So you do not have an extraordinary cost for pulverizing coal in that particular case.

B. Liu: I would like to ask if any panel member would like to comment on the remarks by any other member.

Benarie: Yes, just a brief comment. Measuring costs of taking out substances is not a good measure of particulate separation because as everyone knows if I take out one kilo of substance, it means 99.999% purities cost a lot more than 50%. Secondly, your price is to be compared with flue gas desulfurization. As you know the EPA made a very important 400 page report about the cost comparison of 6 possible methods of flue gas desulfurization. After analyzing all, the cost is between 0.6 and 2.4 cents per kilowatt depending on the method.

Oder: Yes, that is an operating cost figure. Costs for flue gas separation have been estimated at 16 dollars per ton of coal fed into the plant. That is expensive. And furthermore, you have to realize that flue gas desulfurization is tied to the operating factor of the plant. When the plant is down, you are still paying for the installation. But if you are not running, you are not burning clean fuel. I think that is a significant point.

Drehmel: We are entering the area of comparison between different types of control devices and I think it is a very difficult area to consider. Not only can we use the types of measures already mentioned but of course we also use dollars per kilowatt. People want to know what is going into capital equipment. We can also talk about how much it is going to cost the consumer. That is just the economic basis of the comparison. We have talked about power. We might want to talk about resident time or total space and floor space. The ways in which you could compare a device are so diverse from one application to the next that the only really fair way to compare is to take an actual circumstance and say, "I've got this type of aerosol with this type of gas and these types of retrofit problems." Is it a retrofit or I can redesign with some new installation ? What can I do under these circumstances? Look at the actual design. As soon as you start talking in generalities, about watts per CFM, dollars per CFM, you are going to get into trouble.

Liu: I would like to open this up to the floor now.

Baril: Do you envision that this cleaning of the coal takes place at the sight of burning?

Oder: I really didn't intend for that to be coal. I was trying to get into gas-particulate but we have done an extensive study for the Bureau of Mines where we considered the chemical cleaning of coal. This is short of coal conversion such as gasification. This would include, however, physical concleaning and some modest amount of hydro-thermal-chemical operations such as ditioning of fine coal dust. Dry coal cleaning employing pulverizing might best be developed at the power plant location unless economical methods are developed for compacting dry pulverized coal. With this technology coal cleaning could be located remote from the power plant site.

Baril: My second question is, unless you could do it and do it very well to meet the standards, you would not replace the necessity of having to clean the exhaust from a coal burning plant.

Oder: It has been estimated that about 21% of the mid-Western and Eastern U.S. bituminous coal reserves can be brought into compliance at the 1.2 lbs SO$_2$/mm Btu level utilizing physical cleaning technologies with the capability of removing 95% of the mineral sulfur in the coal. Chemical cleaning, with the ability of removing 40% of the organic sulfur, is estimated to have the potential for bringing about 38% of these reserves into compliance. The remaining coals might be brought into compliance by a combination of technologies employing physical & chemical coal cleaning, coal blending and flue gas desulfurization. The extent to which this mix is employed will depend critically upon component economics and upon sulfur emissions standards which must be met. The proper mix of technologies will most likely be determined on a case by case basis.

Billings: It seems to me the next question would be: What is the iron ore benefit to that process? You have half of the weight percent which is iron that you are going to get out of it and you ought to be able to sell that for $10 to $20 per ton or something like that.

Oder: We have not credited any of the processes.

Billings: Is it a major credit?

Oder: Actually the sulfur could be a major credit in Brazil. The first studies by Sergio Trindade in coal cleaning with magnetic separators looked at coals from Brazil. Brazil has a lot of dirty coal but doesn't have much sulfur. Trindade's emphasis was not environmental, but rather to pursue recovery of pyrites because of their sulfur content. The next major application of magnetic separation, interestingly enough, could occur in processing of rich iron ores. And these ores are mined and beneficiated for the purpose of iron ore recovery.

Billings: I don't know what ore goes for but it has to be $10 to $20 a ton. You start adding that up to a hundred million tons a year of iron ore for the 1985 capacity in the U.S., then you start looking at 5 million tons of coal and it sounds like you could meet the iron ore requirement from this beneficiated material.

Oder: It could be, but we didn't consider it.

Billings: Well, if you got a hundred million tons of iron out of that process, it would make it a very attractive way.

Oder: Iron and sulfur could, in principle, be recovered as values in physical coal cleaning. However, I feel that annual recoveries on the order of 100 million tons of iron are out of the question since the 1973 market for coal supplied to electric utilities was itself only 380 million tons. If 21% of this utility coal had been cleaned completely of mineral sulfur using an iron carbonyl process (32 lbs iron carbonyl/ton of coal), then I estimate about one million tons of iron at most would have been available as iron-rich disulfides. While this is not insignificant, it falls far short of the 100 million tons per year quoted by you.

Walston: I agree with what Robin has said - but just to keep us honest and I have vested interest in coal - if we can take out the pyrite, the main goal in the U.S. is SO_2 reduction. But we still will have a particular problem which I am sure everyone is aware of. I just wish to make that comment and also that a couple of times during this workshop my name has been mentioned in connection with some HGMS model work. I want to mention that Professors Richard Treed and William Simons were full collaborative investigators in that.

Melcher: Just a suggestion regarding a question that came up about comparisons among different clean-up techniques. It is a pretty simple matter to improve the efficiency or pressure drop by using something akin to effectiveness factor of a filter that was defined by Chen back in the mid '40's where you used like a negative log of the penetration divided by pressure drop. When you use augmentation of some sort, rather than using pressure drop perhaps you should use total power input. That gives you a clear indication of what you have done. You can always improve efficiency by putting in more layers; of course, it doesn't help you in the pressure drop. And there has been a lot of talk about improving the process this way or that way and it seems reasonable to look at it in terms of the power input.

Oder: That is just the message I was trying to give on the comparison between the magnetic filter bed and the quadrapole. They are diversive approaches to the same problem. One has excellent technical ability from the standpoint of separation but poor performance from the standpoint of pressure drop and total power requirement. The other one has moderate to poor technical performance, high power requirements but no pressure drop problem. I am saying that you have to look at the total device within the process context.

Melcher: In the magnetic separation in both the two types, the bed and the quadrapole design, you end up loosing some of the coal into the slip stream of the coal off the quadrapole or you deposit some in your filter bed. I think Billings reported deficiencies with magnet off as high as 50%. You can't tolerate a 50% loss to your coal if you are going to run that data. I can see it as a clean-up device where you would like to get 50%, but if you are going to separate, you don't want that to occur. How much do you think you can tolerate in terms of loss of coal, for instance?

Oder: I will tell you what can be achieved in commercial practice for kaolin, then I will tell you what we estimated for coal. Again I emphasize that coal is a calculation and an estimate where problems associated with fine clay inclusion, have been ignored. For the processing of kaolin clay which is a wet mineral beneficiation, where 90% of the particles are finer than 2 microns in size a mineral impurity is removed that has a bimodal distribution, 1/4 micron and 8 micron in particle size. The magnetic process operates in a 97% kaolin clay yield. That is one of the main reasons that the technology has won out over such alternative technologies as differential flocculation and flotation. Now the coal that we used for the Bureau of Mines survey is the same coal that has been considered in the paper presented here. It is a coal that is chosen on the basis of several levels (1) the organic sulfur level was low enough for the conceptual study such that if you removed most of the pyrites sulfur you would make a coal that comes into compliance (that seems to be significant) and (2) we chose a coal that cannot be cleaned with a present day cleaning circuit. We chose a coal that had a particle size distribution that corresponds very closely to 200 mesh grind. The feeling was if we grind to 200 mesh, then we will do a very good job of liberation. Given that size distribution, we can estimate the liberation in 200 mesh grind.

On the basis of that estimate, the HGMS technology in the example that I gave should receive about a 95% Btu recovery. The quadrapole I think had maybe 80 to 85% Btu recovery. It is in the paper. There is obviously a liberation problem; you will have to pulverize the coal to eliminate the mineral matter.

Melcher: You are saying then that perhaps something like 10% of your coal is lost.

Oder: Look, no one has done this. I am guessing that it is 10%. If you have problems of coalescense of the particles like I have been listening to here, the loss could be a great deal more. If one pyrite particle (even though it is ground to liberation) has 2 or 3 coal particles struck to it because of electrical attraction, when you take out the pyrite, you might take out a lot of coal. The estimate is based strictly on a simplified, theoretical calculation.

Nichols: I thought I would make one comment about your magnetic quenching sparking in your electrostatic circuitry. Anyone who is going to do research in this area should consider the fact that the energy is stored in the electrode system rather than at the power supply. So your magnetic quencher has to be distributed throughout the whole corona wire conduction electrode system. The majority of energy that goes into this part is stored energy that is in the distributive capacitance of the system.

Oder: Yes, I wouldn't know how to do it. I would suggest either couple it inductively or put in a permanent magnet.

Nichols: Yes, that is one area they might do research in. It has to be in a box.

Friedlander: First, last night I gave it some thought and I guess this existing system has some merit to it, because it quenches quickly, it localizes the arc, but I also would like to consider any method that you always quench after you stop the arc. In magnetic method you use nothing to quench unless you stop it. So you still have a problem there. I want to get back to something. My feeling is quite strong about the many technologies of the agglomerations sticking together. It is really important in two ways, (1) To avoid where you don't want it and (2) To use it in seeding. Two years ago I read about water-based technology of seeding water with iron oxide to remove bacteria viruses. Apparently it works very well and I have not heard much more about it. The air system,I haven't heard anything about it yet except here. There are some problems that can be solved that way. But I think a lot of work has to be done to understand the system and the technology to make it work. The magnetic technology is a peculiar one. There are not that many particles around today. But what about seeding magnetic particles, then getting that magnetic particle to marry another magnetic particle? It is not the usual agglomeration problems where you have to get a lot of them together. If you don't have high densities, you only need one magnetic particle to remove particles so you only have to marry two particles and not a large number of particles. The time constant may not be that bad.

Liu: I would like to change the topic just slightly. I think I would like to ask Don Reid about his electrostatic augmented filters. I am just wondering: eventually the filter will clog up and you have to replace it, don't you. I just wonder what type of application you think your equipment would have.

Reid: We haven't done any extensive studies of the cleaning techniques. Again, we do know that we can clean it back to the original pressure drops, which indicates that we have a clean bed. It has a high loading capacity. We have operated some of these filters for 60 hours and only had 0.1 inch increase in pressure drop. Of course the length of operation time before cleaning is going to be a function of concentration of the aerosol.

Liu: So you are intending to use this as a clean up filter. You wash it after it gets clogged.

Reid: You have to clean it. It is not one you just throw away.

Liu: I see.

Billings: I wonder if Professor Löffler would give us the benefit of his thinking about the strengths of aggregates or agglomerate particles. What the forces are, what the orders of magnitude are, that kind of thing. Would that be a fair question?

Löffler: I think it would be a topic for another meeting if you talk about agglomeration processes between various particles. I could only give a very brief review of the main mechanisms. We could list all the forces which are possible, we can go back to non-material binding, such as Van der Waals forces, electrostatic forces, or any other dipole interaction. We have worked with a material which is like a liquid (water or something like water) where there are chemical forces, surface tension forces and you might have some solid which is crystallized. Those are some of the mechanisms. As far as magnitudes: The highest of course is a solid which might also be an effect of surface diffusion. The next range is the liquid which also acts over large distances. The magnitude below that you get the van der Waals forces. The order below that, electrostatic forces. As long as the particles are in direct contact the distance must be in the range of several angstrom units. But electrostatic forces are long range forces. To give you an idea about the orders of magnitude: For a 1 μm particle, the van der Waals adhesion force, to a surface would be one million times the gravity. That should give you an estimate on the amount of energy needed to deagglomerate. This means that it is normally impossible to deagglomerate very fine particles only by shear forces in a gas flow. It is easy through the use of impact forces. This way you can deagglomerate particles. We found that in deagglomeration, impact onto a hard surface, you need velocities approaching 200 meters/second.

Oder: Professor Friedlander asked if anything was occurring in the seeding technology for wet processing with magnetic separators and the answer is "yes". It bears on what you are saying. There is a great deal proceeding from the standpoint of industrial development for processing the effluent water from municipal and industrial complexes and it is a technology that utilizes the seeding technology. Coagules are formed in a normal flocculation chamber where you change the pH so as to make the particles coagulate and form on the surface of magnetic iron oxide particles which are about 20 microns in diameter. Now what has been found is that the magnetic capture force for these coagules in high velocity flow through that packed bed is so great that it is shear force that is breaking up the coagules. The process rate through the magnet is limited by the breakup of the coagules and not by the availability of a magnetic force to hold the seed material on the capture surface. In that case we have a seed material that is rough in shape, about 20 microns in diameter, and that can capture four or five times its volume in entrained materials. Studies have been made of capture of iron oxide particles, oil agglomerates and even bacterial systems which are the products of tertiary and secondary sewage processing. You can capture and hole these up to a point, but as the flow velocity increases eventually viscous forces will break the coagules.

Billings: I think that there seems to be a general concern, both on the gas side and the liquid side, to attempt to understand something that is pretty fundamental. That is, what is the force of adhesion between two fibers, similar and dissimilar, and how does that compare what those forces are? First, where do they come from? Second, how does one speculate about the order of magnitude? Then you can begin to compare those forces between the particles with the other forces that come into play, like the fluid mechanical forces. The forces of adhesion between particles in gases at normal temperatures and pressures are enormous. They are very much higher than in liquids. It becomes very difficult to go from one fluid to the other because the adhesion forces dominate in the gas side.

Oder: Is that because of double layering in the liquid? Do you have shielding of the charge in some parts?

Billings: From an engineer's point of view, the adhesion force of the particles in a gas without the Van der Waals, without electrical, without sintering, without actual contacts is primarily the surface tension of water. It is essentially a humidity effect. It is demonstrated very easily by varying the humidity and examining the force between the particle and the surface. The point is that tension force is higher than anything else. The adhesion force is of the order of 5 dynes. The weight of the particle is on the order of 10^{-9} dynes. The drag force = $3\pi\mu V_d$ is of the order of 10^{-8} times the velocity. So the velocity has to be of the order of 10^8 meters/sec in order to blow that particle off the surface.

Oder: Question. Do you know of, or does any one know of, dispersing materials to be used in dry systems? Wet particulates can be stabilized in fluids with detergents effectively. One adds polyphosphates so that the particulates become stabilized by charge repulsion. Are there equivalent materials for use with dry material to promote stability in separation in particles?

Liu: An answer to that is no. A number of chemical warfare people have tried for many years to stabilize the aerosols so that they would not flocculate and coagulate. I think the efforts have all been negative. You do not have an equivalent double layer situation in a gas system. When they come together, particles simply stick. I think one of the biggest problems with gas separation is you have to deal with very high concentrations of particulate in order to reduce the volume of gas in the process. That is conducive to coagulation. Coagulation has to be looked at very carefully before one can say anything about the economic feasibility in magnetic separation.

Scott: I just want to return to the electrostatic filter. Question: It is not my field, but what is the physics or mechanism that allows this particular process to occur with much lower pressure drops. I'm particularly intrigued by one of the slides where you had low pressure drop, then you turned off the electric field and the pressure drop went up; then, magically, you turned on the electric field and I think it dropped again. I wonder if I misinterpreted that or could someone give me a little bit more physical interpretation of that part.

Ariman: Dennis, would you please make some comments on the last part of the question?

Helfritch: When the electric field is turned off, we do know we have briefly a precipitated section that the aerosol has to pass through before it is filtered by the fabric. Some of it is taken out. Second, when dust deposits on the fabric in the charged state, it deposits in a different way, more porously and openly. When the charge is turned off, the dust then can deposit in its normal fashion, which is a fairly tightly packed structure with a lower permeability. We clean the fabric periodically. The entire unit is cleaned every 5 minutes.

Ariman: Let me add a couple of things to his comments. We do have a more porous dust cake when we have an electric charge on the particles. In the pictures you will see just the particles are charged. We do have charges on particles, and it looks like they keep their charge at least for a while depending on the filter. In bench scale experiments once one particle lands on the filter the next one because of the charge of the first particle lands somewhere else on the filter. However, the second one would land next to the first particle in the case of non-charged particles. Therefore for the charged case there is a nonuniform distribution of the landing of particles on that particular filter. As I indicated earlier there are little hills, valleys, and bridges. If you are not careful in carrying the dust cake to the microscope, you can damage the cake structure. It is very fluffy and porous. We hope to shed more light into it in the near future.

Liu: I would like to suggest one of the more important mechanisms might be due to the poly dispersity of the dust. If you do not have an electrostatic charge, then the fine particles would have a tendency of filling the pores because this is where it can be collected by interception. Electrostatically charged aerosols tend to be collected at the entrance of those pores, probably keeping those pores open so that the fluid can go through.

Cooper: One of the things that struck me was decrease in the flow resistance of the dust cake as humidity increased and adhesive forces which might tend to pack it tighter and electrical forces which may decrease by humidity. Those results are surprising. The humidity is allowing us to build up a more open cake simply because once we get a particle striking another, it stays in a forward position rather than going into a backward position.

Ariman: First of all, let me clarify one thing. The results provided here were not for the charged cakes. They were obtained by the use of single miniature bag testing devices. We really tried to compare our bench scale testing which uses flat fabric with a regular single bag, but in a small scale. So we did not have any charge. Also we did have greater complexity when we had electric charge because of the interaction between humidity and the electric field. Now I'll go back to your first question - how come the humidity was causing a decrease in the pressure drop? It is my understanding, that we do have a change in the adhesive forces between the particles when relative humidity varies. Also the same thing applies for the forces between the particles and the fiber filter surface. It looks like there is an increase in agglomeration process of the particles before the landing on the filter. Instead of having very tightly packed dust cakes we have quite porous ones.

Penny: There is one factor that wasn't mentioned in connection with humidity and that is the question of resistivity. For the information of the porous cake the surface has to be in equipotential. If the surface is equipotential then the field will concentrate on any projection. It turns out that requires a certain amount of conductivity for the surface to be in equipotential. So it seems to me that when we mention humidity, we ought to mention the kind of material and how its resistivity varies from humidity. In other words, it is not sufficient to have charged particles in that cake - you really want that charge conducted. If you do it right you build up these particles into almost like fiber like clouds. But that requires high enough conductivity.

Ariman: I agree with Professor Penney. Resistivity is another area that I was going to suggest as equally important with relative humidity. I think we should look into the resistivity very, very closely and for bench scale testing, we should find a way to measure the resistivity of each dust particle we have been using. We should also try to incorporate resistivity into the mathematical model of filtration in an acceptable way.

Löffler: What is needed is to go into more detail, not in theory but in experiment, to get more information on the material qualities. All the effects that were observed rely on very specific materials and conditions. For example, we did measurements on low humidity using a very similar apparatus as you did and found that in this case the particles formed a cake. There was a continuous variation in the permeability of the cake. That means the cake was compressed. There was an increasing pressure drop across this cake and this could be theorized by going for a short moment with a higher velocity and with a lower velocity. You would obtain a lower permeability. I would speculate that in your case one of the possible reasons is with increasing humidity you get a liquid which is at the contact point and is without a higher adhesion force in stronger agglomerates. This type of cake perhaps does not reduce permeability.

Dr. Friedlander: Basically this research turns out to strongly involve the property of solid materials. On the other hand, those of us at the materials research laboratories, are told that any work that considers such things can not be considered to have anything to do with money that is given to the study of materials which is also contradictory because we end up right at this point finding that where the lack of knowledge is extremely strong is in the understanding of the solid materials involved. The NSF says that this is not so. But yet it is interesting to find that the interaction between these two may be extremely important. It should be encouraged rather than discouraged.

Liu: Dr. Ojalvo, I would like to hear your comments on that.

Ojalvo: I certainly agree and it takes workshops like this to bring this sort of thing out. Eventually this reaches the people at the top of the NSF ranking and decision-making and it has been argued strongly enough, forcefully enough, and a number of times. It is good this sort of work can be brought together.

Billings: It seems like Dr. Friedlander's comments should be modified in the following way. That we really aren't interested in the bulk properties of material. To a great extent, these are inconsequential. We really are interested in interfacial property. And that has nothing to do or very little to do with the material science area.

Friedlander: Surface studies are the big area - Purdue University does surface study and get well paid. This is precisely the interaction of surface study on the solid material, the oxide. And here we find a different viewpoint but exactly the same property, when you get down to real basics, we are talking about the same thing they are. Maybe they know a lot more than we do except maybe it doesn't apply to this function. We are saying the same thing but for a different situation.

Lamb: We have reviewed reports by Löffler on a phenomenon which he has reported reducing the pressure drops by a factor of six. The important thing is that he describes this and nobody knows why it is happening. I am somewhat discouraged that he lists several papers on the theory of fiber filtration that always come to deal with a single fiber being struck by a single particle. They always deal with conditions that can be described as fictitious. The situation is one in which we are coming to know very precisely whatever happens. The filter is always clean and we never see what happens when the second particle comes along and is faced with a different situation. Perhaps it is time that we devoted efforts to this very messy situation.

Ariman: I agree with you, Dr. Lamb, that is one of the major points that I have tried to stress in my presentation. The time has come to go from this single fiber, single particle approach step-by-step to the real filtration with close coordination of experimental work.

Liu: The time is getting late. Some people have to leave for the airport so I think we should close at this point. So before we do that I would like to ask Dr. Ojalvo and Dr. Drehmel if they have some closing remarks.

Drehmel: It seems to me that one thing that we really haven't addressed is how good our theories are. Do we believe them and what happens when that second particle comes in? I agree with Dr. Lamb - there are a lot of other theories that can be proposed. So far as comments on this workshop are concerned, I think that we owe a great deal of thanks to Dr. Ariman for organizing this conference and getting the facilities for us. I think the attention to details that he has given us on this workshop has just been fantastic.

Ojalvo: I would like first to make some general remarks about the National Science Foundation (NSF) and then conclude with information about my program and its relationship to this workshop.

The basic purpose of NSF is to initiate and support scientific research (including social science and engineering) and programs with scientific research potential - such as science education. To accomplish this purpose NSF is organized into seven directorates, each headed by and Assistant Director, under the Director of the NSF. The first three directorates are concerned with basic research - Mathematical and Physical Sciences and Engineering; Astronomical, Atmospheric, Earth and Ocean Sciences; and Biological, Behavioral and Social Sciences. The other four are: Science Education; Research Applied to National Needs (RANN); Scientific, Technological and International Affairs; and finally a Directorate for Administration. The first three directorates (basic research) are the heart of NSF and account for 70% to 75% of its total budget.

There are five divisions in the Mathematical and Physical Sciences and Engineering Directorate - Physics; Chemistry; Mathematical and Computer Sciences; Materials Research; and Engineering. The Engineering Division consists of three sections: Electrical Sciences and Analysis; Engineering Mechanics; and the section with which I am associated - Engineering Chemistry and Energetics. The program which I direct, Solid and Particulate Processing, is one of five in this section. The other four are Heat Transfer; Engineering Energetics; Thermodynamics and Mass Transfer; and Chemical Processes.

The four major areas in my program are: Particle and Particle System Characterization; Particle Generation and Size Modification; Interfacial and Colloidal Phenomena; and finally, the area of interest in this workshop - Particle Processing and Separation.

The program is quite new having been initiated two years ago. One of the first activities was to hold workshops. There was much interest from mineral engineers, so the first workshop was on Mineral Processing. Then, the Fine Particle Society conducted a workshop on Particle Technology. One of the conclusions of that workshop was that particulate - gas separation was an area that could be explored in a separate workshop by itself. So here we are doing just that!

I want to congratulate Dr. Ariman for conducting a very fine and productive workshop. I feel that quite a bit has been accomplished. I would like to have the Proceedings include all the papers, invited lectures, discussions and comments presented here so that we at NSF (more specifically the Solid and Particulate Processing Program) and others in the technical community can use it as guidance to focus our efforts on the promising research opportunities in particulate - gas separation.

Ariman: Thank you, Dr. Drehmel and Dr. Ojalvo, for the nice comments you have made concerning the workshop. First of all, I would like to thank our distinguished lecturers, Professor Löffler and Dr. Benarie. They were kind enough to come from Europe on such a short notice. I also would like to thank the lecturers and the designated discussors for their fine contributions and certainly the important contributions of all the other participants through open discussions.

It is proper also to express my appreciation and thanks for the great help to the members of the organizing committee, Professors Ben Liu and K. T. Yang. Also special thanks to our sponsors NSF, EPA and host institution University of Notre Dame. Without their help, there was no possibility of having this workshop. Very special thanks to Dr. Ojalvo of NSF and Dr. Drehmel of the EPA, for their continuous support from the beginning and for their wise advice and suggestions. So, on behalf of the organizing committee and the University of Notre Dame, we are very grateful to them. I hope it was a useful workshop for every participant and I would like to congratulate all participants for their very valuable contributions.